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# SLUDGE PROCESSING FOR COMBINED PHYSICAL-CHEMICAL-BIOLOGICAL SLUDGES

D. S. Parker, et al

Central Contra Costa Sanitary District
Walnut Creek, California

July 1973

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# Sludge Processing For Combined Physical-Chemical-Biological Sludges

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16. Abstract

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# SLUDGE PROCESSING FOR COMBINED PHYSICAL-CHEMICAL-BIOLOGICAL SLUDGES

by

D. S. Parker
F. J. Zadick
K. E. Train

**For the**

**Central Contra Costa Sanitary District**
**Walnut Creek, California 94596**

Grant No. R801445
**Project No. 17080 FSF**

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## ABSTRACT

Full scale combined sludge generation from a treatment sequence consisting of lime clarification, nitrification and denitrification was studied. Pilot scale studies were conducted to wet-classify and dewater the combined sludges by means of two-stage solid bowl centrifugation. Anaerobic digestion of first stage centrate was also studied on a pilot scale.

Predicted and measured sludge production agreed well using four coagulation modes in the chemical primary.

First stage centrifugation (wet classification), over a wide pH range, achieved high capture of calcium carbonate in relatively dry (42 to 57 percent total solids)cakes as well as good rejection of magnesium, phosphorus, and iron compounds in the centrates.

At pH 11.0 or below, second stage centrifugation dewatered first stage centrate to produce 18 percent total solids cakes, with 80 percent solids recoveries. Dewatering deteriorated at a pH greater than 11.0.

Anaerobic stabilization of thickened and unthickened first stage centrates showed volatile matter destructions of over 40 percent, high methane-content gas, and substantial increases in soluble magnesium and alkalinity during digestion.



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# SECTION I

# CONCLUSIONS

1. Combined sludge generated in a chemical primary sedimentation tank where lime is the coagulant can be accurately estimated from changes in quality constituents across the process and knowledge of chemical doses.

2. Considerable sludge thickening takes place in the primary sedimentation tank. Only the first half of the rectangular tank, however, is effective for thickening. Thicker sludge is obtained at pH 11.0 and 11.5 than at pH 10.2.

3. Combined sludge can be effectively classified with a solid bowl centrifuge. For a flocculation pH of 11.0 and 11.5 approximately 90 percent of the calcium carbonate fed to the machine was recovered in the cake. At the same time, 50 to 75 percent of the other constituents were rejected in the centrate. The calcium carbonate-rich cake was simultaneously dewatered to a total solids content of greater than 50 percent. At pH 10.2, the calcium carbonate recovery was 75 percent while the rejection of the other materials was 60 percent. Cake dryness at pH 10.2 was 42 to 49 percent total solids.

4. Centrate from the first stage classification centrifuge can be dewatered in a second stage solid bowl centrifuge. Anionic polymer addition is required at a dose of at least two lb per ton of dry solids. The pH of operation affects the cake dryness obtained. Below pH 11.0, total solids in the cake of 17 to 18 percent can be obtained. Above pH 11.0 dewatering deteriorates, and a median cake total solids of 12 percent was produced.

5. Anaerobic digestion can stabilize first stage centrate, but volatile matter reduction is less for centrate processing than for conventional raw sludge processing. Further, virtually all of the magnesium hydroxide fed to the digester is solubilized due to the pH of digester operation (pH 7.0 to 8.0). The supernatant from digestion, if returned to the process, would cause an increase in the lime dose required because of recycled alkalinity and magnesium. Calcium carbonate and hydroxyapatite are not dissolved in the digestion process to a significant degree.

6. Combined sludge can be stabilized by chlorination treatment. However, supernatant from a chlorination stabilization step cannot be returned to the process, as the total dissolved solids of the reclaimed water will be increased.

7. Two feasible solids disposal alternatives have been developed for combined sludge processing. One involves incineration of the waste solids with separate recalcination of the calcium carbonate. The other solids disposal alternative involves disposal of the calcium carbonate in landfill with digestion of the waste solids. The incineration/recalcination alternative is preferable, since chemical costs are significantly less and far less solids are produced for ultimate disposal.

## SECTION II

## RECOMMENDATIONS

1. One of the most pressing needs of designers of advanced waste treatment systems is for meaningful design information on solids disposal systems. While this report presents information useful for some of the advanced waste treatment systems, there is a large number of systems for which there is no data at all. An intensive research effort is required to fill the gaps that exist.

2. It is recommended that future investigations of processing combined sludges include the following:

   a. Material balance studies on heavy metals across each step in the solids processing scheme.

   b. Metal oxide material balances across each solids processing step where incineration is employed in the processing scheme.

   c. Additional studies on dry classification of ash from recalcining furnaces.

   d. Effects of temperature on thickening of chemical combined sludges.

   e. Effect of recycled material (inerts, heavy metals, etc.) on an integrated solids two-stage centrifugation, recalcination and incineration.

   f. Studies of chemical sludge stabilization by chlorination with subsequent land disposal.

3. It is recommended that solid bowl centrifuges employed in wet classification and dewatering of combined physical-chemical-biological sludges should be designed to include the following features:

   a. The capability of stepwise variation in the centrifuge force from 1100 to 3100 times the force of gravity.

   b. The capability of stepwise variation of the bowl pond setting for liquid discharge levels.

   c. The capability of continuous variation of auger speed relative to bowl speed.

d. The capability of internal polymer introduction to the bowl.

4. Anaerobic digestion of first stage centrate from centrifuging combined sludges should be designed to operate with feed solids of at least 3 percent volatile solids. Also, solids retention time should be close to 20 days to attain a 40 percent volatile matter reduction.

## SECTION III

## INTRODUCTION

Increasingly, waste treatment systems are being designed to produce effluents of substantially higher quality than can be obtained by conventional primary or secondary processes. Most commonly, these new "advanced waste treatment" systems are oriented toward the removal of nutrients in order to prevent over-fertilization of receiving waters. These advanced treatment systems commonly remove either phosphorus or nitrogen, or both. Other systems have been developed for enhanced heavy metals removal.

The removal of phosphorus normally entails precipitation by a metal salt, which may be iron (ferric or ferrous), aluminum (alum or sodium aluminate), or calcium (lime). These coagulants may be used in the process for other purposes, such as heavy metals removal. The use of the metal salt substantially increases the mass of sludge which must be removed by either primary or secondary treatment. Further, the dewatering and solids processing characteristics of the resulting sludges are substantially different from either primary-generated raw sludge or secondary-generated biological sludges. Since there are very few operational full-scale advanced treatment systems (most results to date are from pilot plants), little information is available on which to base the design of advanced treatment solids processing systems. Nevertheless, wastewater agencies and consultants are faced with the immediate problem of designing full-scale solids processing facilities for a host of advanced treatment systems.

This project involved the study of solids processing for an advanced treatment system, termed the ATTF system. The ATTF system is a tailored sequence of processes specifically developed for the purpose of reclaiming municipal wastewater for industrial use and meeting discharge requirements [1].

The basic ATTF flow diagram is shown on Fig. 3-1. The process couples lime clarification with a combined oxidation-nitrification step. The use of lime in the primary treatment stage removes much of the organic carbon load from the oxidation-nitrification stage, thus allowing stable oxidation of ammonia to nitrate. In addition to the removal of organic matter, the lime enhances the removal of phosphorus, heavy metals (toxic materials) and viruses. Other advanced treatment systems have used lime in tertiary treatment after conventional secondary treatment, but by bringing lime addition forward to the initial stage of treatment, the load on the treatment processes which follow is reduced, and operation is stabilized. Biological denitrification follows nitrification, converting nitrate to nitrogen gas.

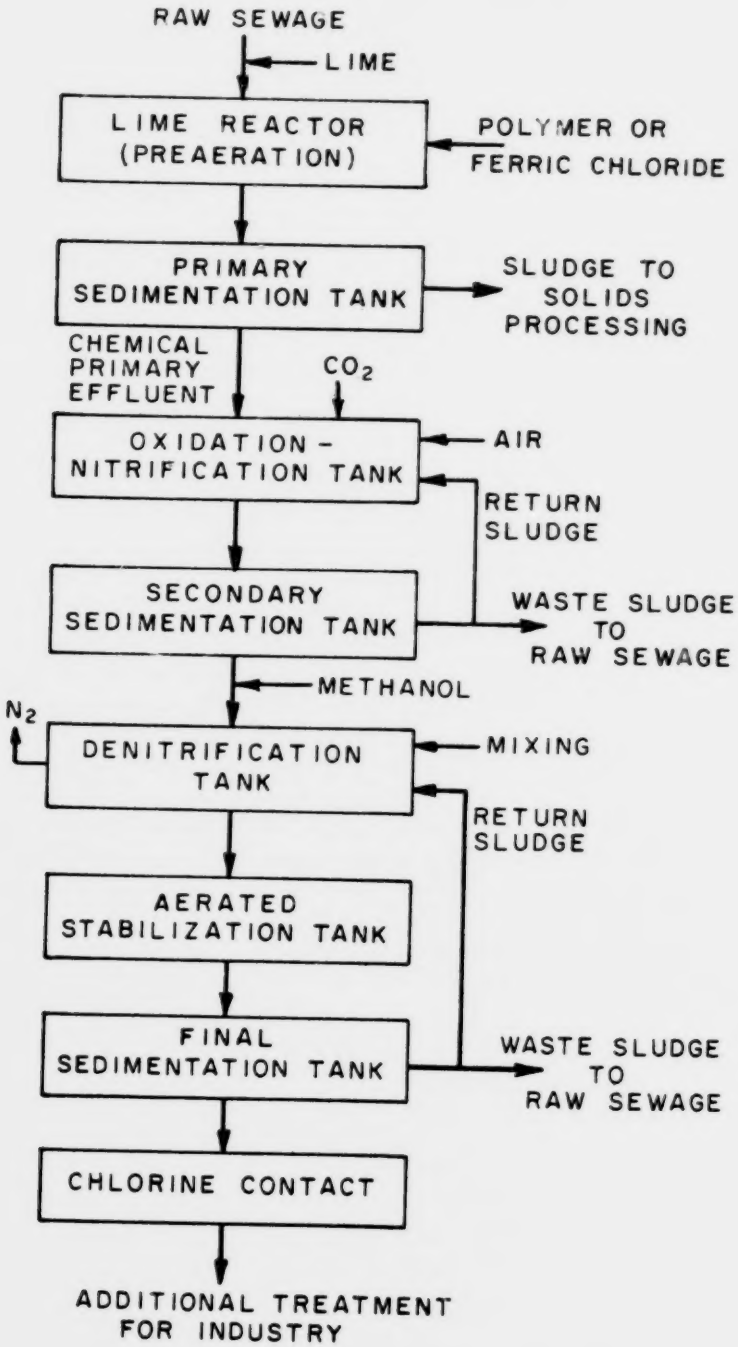



Figure 3-1

ATTF System

Although not shown in the diagram, multi-media filtration will be provided after denitrification to polish the flow stream before distribution to industry or discharge.

The ATTF system produces a combined physical-chemical-biological sludge. Raw sewage solids, calcium phosphate (hydroxyapatite), magnesium hydroxide, and calcium carbonate are precipitated in the primary process, together with waste biological sludge from the nitrification system and ferric hydroxide when iron salts are used. The primary underflow thus contains all of the sludge produced in the system.

The specific purpose of this project was the development of useful design and operating information for the 30 mgd Central Contra Costa Sanitary District (CCCSD) Reclamation Plant, which will be under construction in early 1973. As a result, avenues of investigation had a more practical orientation than a pure research project would. The results of this research however, will have application wherever lime is used the primary stage of treatment. Examples are:

1. Lime precipitation in the primary for phosphorus removal, followed by biological treatment by activated sludge for BOD removal[2,3,4].

2. Lime precipitation in the primary for phosphorus removal and organics reduction, followed by activated sludge for nitrification[5].

3. Lime precipitation in the primary, followed by filtration and activated carbon adsorption[6,22].

4. Lime precipitation in the primary for improved organics, grease, and heavy metals removal, followed by ocean disposal[7].

The first two categories listed will produce sludges nearly identical to the ATTF system sludge; the latter two categories will produce sludges with similar characteristics.

Sludge processing investigations for the ATTF system were oriented towards segregating recalcinable calcium carbonate from the other sludge constituents, which would build up as inerts in the system if not removed before the lime sludge was recalcined and reused. Recycle of reusable lime is economically desirable because it reduces the new lime requirement of the process and reduces the quantity of ash destined for ultimate disposal. Further, much of the cost of recalcination would have to be spent in any case as a sludge processing cost.

The first stage (30 mgd) of the CCCSD Water Reclamation Plant will employ the solids processing system shown in Fig. 3-2. A two-stage centrifuge system will be used to classify the sludge into two major fractions. The first fraction will contain most of the recyclable lime sludge. The second fraction will contain most of the magnesium, phosphorus and organics. The first stage solid bowl centrifuge will be run at high throughput rate and without polymer (deliberate inefficiency) to classify the heavy calcium carbonate into the cake and put most of the lighter organics and chemical inerts into the centrate. The first stage cake will be recalcined and returned to the process, while the first stage centrate will be conveyed to the second stage centrifuge. The second stage centrifuge will be run at low throughput and with polymer for maximum solids capture. Second stage cake will be incinerated prior to disposal, and second stage centrate will be returned to the raw sewage.

A major goal of this investigation was the definition of the effect of changes in process variables on the operation of each element in the solids processing system. Particular attention was paid to sludge thickening in the primary and to the operation of both stages of centrifugation.

Since the initial construction constitutes only one-quarter of the ultimate plant capacity (and the waste sludge furnace can be a recalcine furnace in a future stage) alternatives for first-stage centrate disposal system were investigated. A schematic listing the centrate processing alternatives is shown on Fig. 3-3. Primary effort was directed to centrifugation (as mentioned before) and anaerobic digestion as the chief processing alternatives.

The digestion process was carried out on a bench scale with first-stage centrate as feed, thickened to two different solids levels. The effect of solids retention time on digestion performance was studied by simultaneously running four centrate digesters at several solids retention times. Also a control digester was run on conventional sludge at a specific solids retention time. The digesters were well mixed so that hydraulic and solids retention times were equal. The digested sludges were withdrawn daily and portions were subjected to odor and drying tests.

Some limited studies on vacuum filtration were done on the first stage centrate. Also, an electrolytic chemical oxidation processing scheme was investigated briefly.

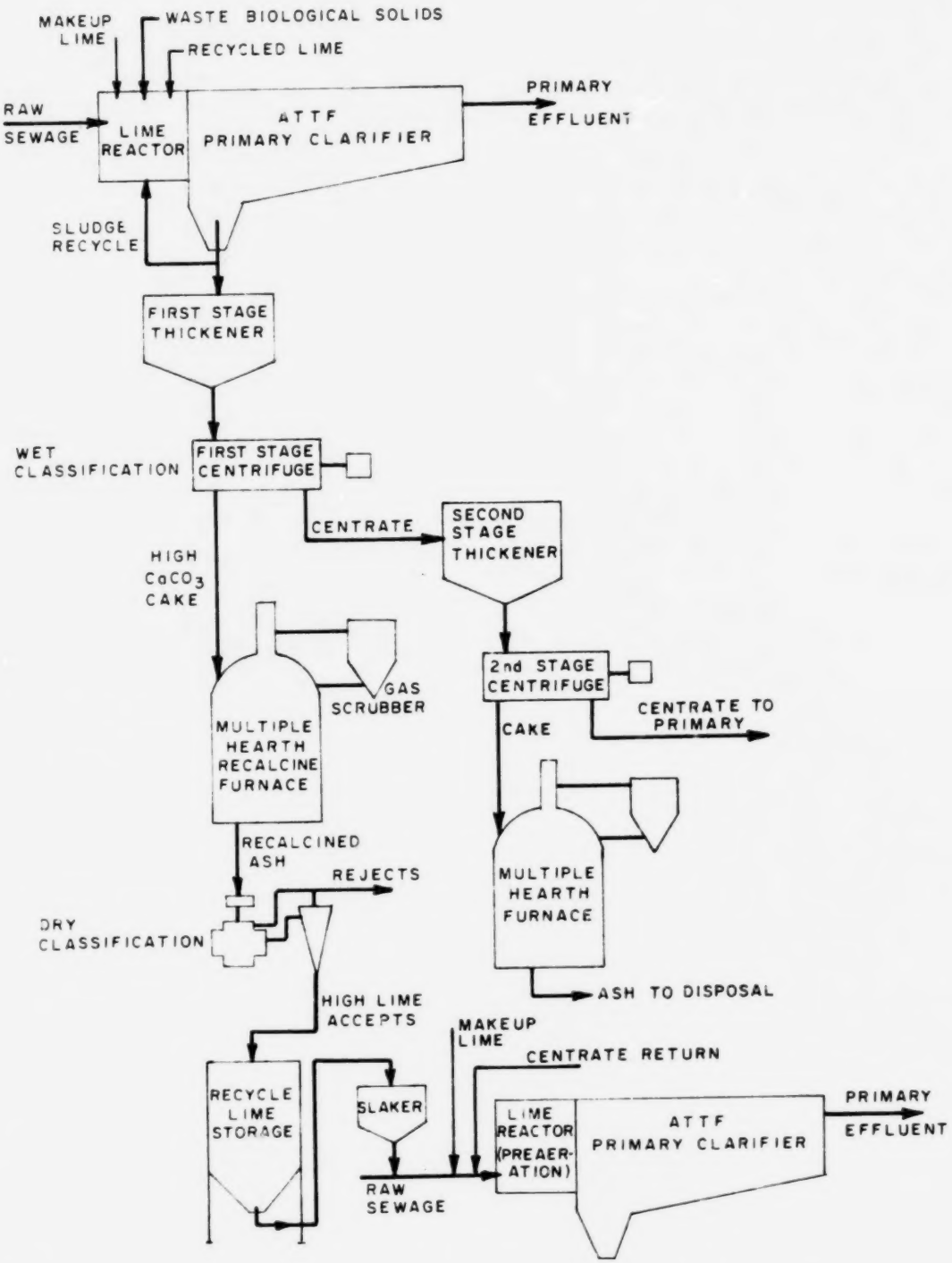



Figure 3-2 ATTF Solids Processing System

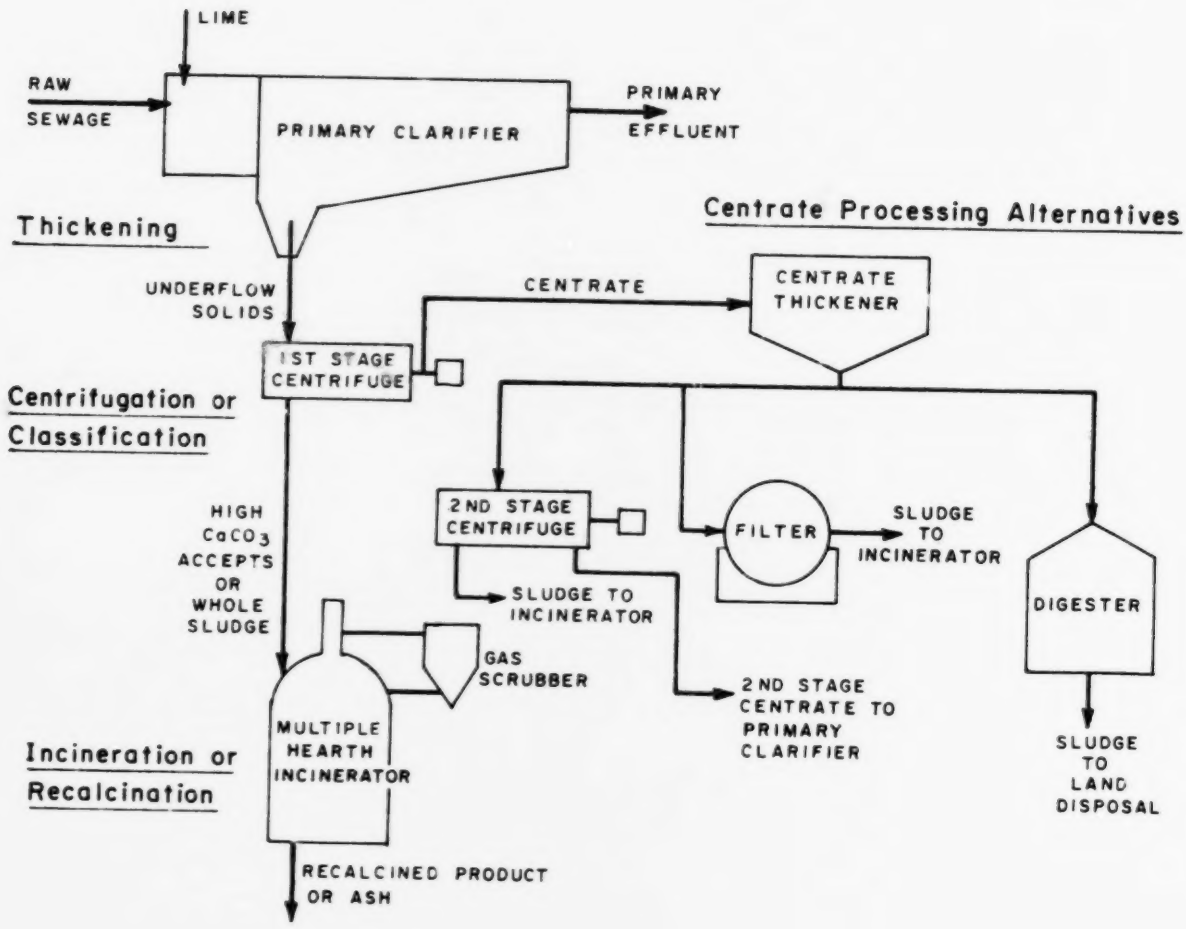



Figure 3-3

Solids Processing - Lime Recovery Mode

## SECTION IV

## ADVANCED TREATMENT TEST FACILITY

The construction and operation of the Advanced Treatment Test Facility by CCCSD is one of a series of steps leading to the design and construction of a 30-mgd water reclamation plant to be completed in 1975. The first significant step towards water reclamation was taken in December 1969, when CCCSD and Contra Costa County Water District (CCCWD) signed a memorandum of understanding which paved the way for a joint effort which will culminate in the recycling of treated wastewater to industries in Contra Costa County. The cooperative program was outlined which included a water reclamation feasibility study, a solid waste disposal investigation, a sampling and analysis program, and a pilot plant demonstration program.

In February 1970, a Class IV research and demonstration grant was issued by the Federal Water Pollution Control Administration (now Environmental Protection Agency) to carry out a two-year on-site demonstration that renovated water could be produced of sufficient quality to replace industry's present raw water supply for most uses. Concurrent with this research work, CCCSD's engineering consultants (Brown and Caldwell) completed a preliminary design report for wastewater treatment expansion and modifications. The treatment scheme recommended for the ultimate plant consisted of lime coagulation-sedimentation, nitrification and denitrification. This process sequence later became known as the ATTF system.

In April 1972, CCCSD completed contract negotiations with CCCWD for sale of the reclaimed water. The water contract set quality limits, among which were 1 mg/l total phosphorus and 5 mg/l total nitrogen. A maximum value for total dissolved solids was established at 375 mg/l greater than industry's present water source, the Contra Costa Canal. The phosphorus limit of 1.0 mg/l as P was set to prevent calcium phosphate scaling of heat exchangers[8]. The nitrogen limit of 5 mg/l was aimed at preventing algal growths on distribution structures and open recirculating cooling towers.

The original CCCSD-CCCWD pilot plant demonstration program had not been designed to evaluate the ATTF process sequence. In separate systems both lime treatment of raw sewage and nitrification were evaluated, but not as parts of the sequence proposed in the ATTF design. Denitrification was attempted late in the pilot program after Brown and Caldwell had proposed the ATTF sequence of processes. However, stable nitrification and denitrification were never successfully demonstrated on a pilot scale. This led regulatory agencies to question the reliability of the proposed ATTF system. In order to estabish the credibility of the ATTF system and generate design information, CCCSD undertook the construction of a full-scale Advanced Treatment Test Facility. During the initial construction of the

ATTF, CCCSD was not assisted by either state or federal research grants. The test facility was placed in operation in late November 1971.

As an integral part of the ATTF system, solids processing had to be evaluated to produce both a feasible processing scheme and design information for the selected processing system. Studies were conducted on the centrifugation of the primary lime sludge to determine the feasibility of classifying the calcium carbonate from the sludge in order to recover lime solids. These studies provided useful design information, but CCCSD sought to undertake a further reasearch effort to obtain larger scale centrifuge performance data as well as to evaluate anaerobic digestion of lime sludges. To finance this effort, CCCSD sougnt research grants for funding the "Combined Sludge Processing Project". Both the State of California and Environmental Protection Agency provided funds for the project. Experimentation began in the summer of 1972.

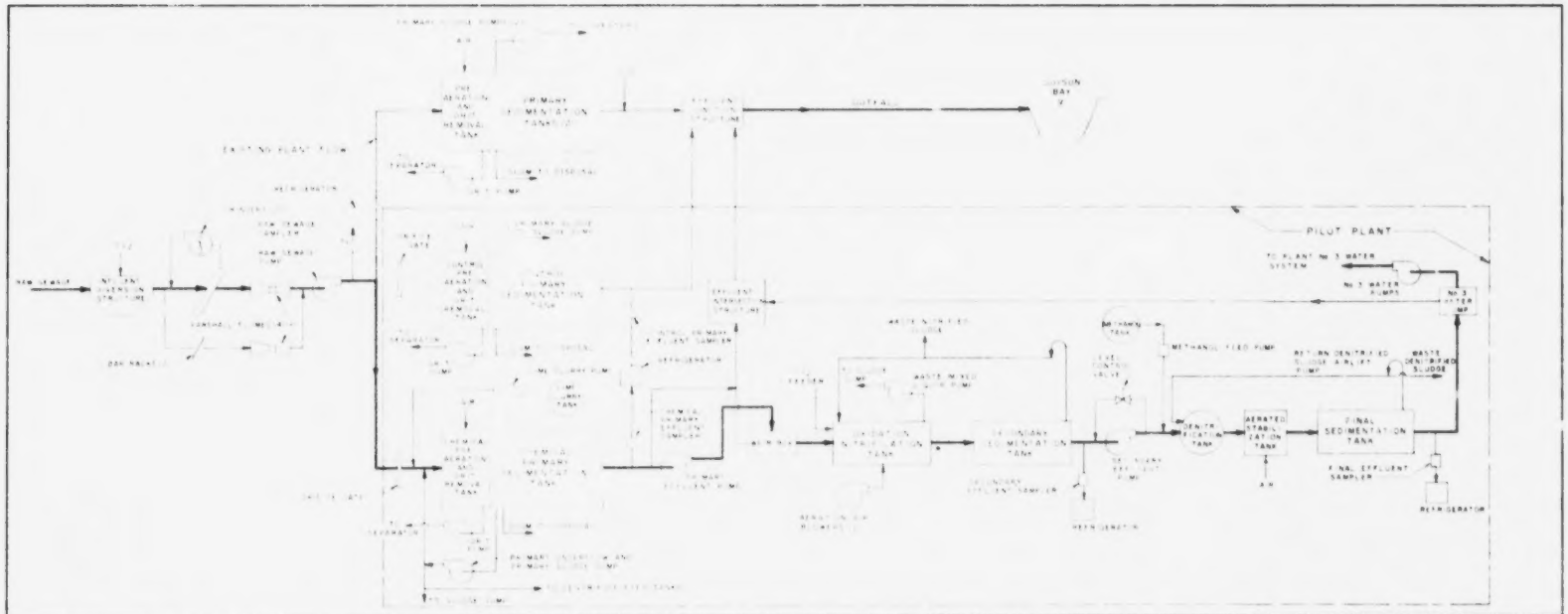
This report deals with the results of the entire research effort on solids processing, including that conducted prior to commencement of grants. Solids processing equipment used in the study is described in this section. Also described are the analytical methods employed in the analysis of solids streams generated in the ATTF process.

## SOLIDS PROCESSING EQUIPMENT

Solids processing consists of sludge generation and handling as well as sludge treatment. The treatment schemes tested in this project dealt with both classification and stabilization of chemical sludges.

### Primary Solids Generation and Handling

The chemical primary operation begins with the addition of lime slurry, ferric chloride, and recirculated primary sludge to the raw sewage flow stream. The next step, preaeration, serves as both a flocculation and grit removal step. Flocculated solids are settled out as chemical primary sludge in the primary sedimentation tank. The settled sludge is removed from the primary tank by a variable speed pump through a timer-actuated valve which regulates the relative proportion of wasted to recirculated sludge. Wasted sludge is pumped to a digester together with the conventional raw sludge from the parallel control primary tank. Recirculated sludge is returned to the preaeration chamber to maintain the high solids level necessary for efficient flocculation. A flow diagram and design data for the chemical primary and other features of the ATTF system are shown in Fig. 4-1.




PILOT PLANT DESIGN DATA

EXISTING TREATMENT PLANT

PLANT LOADING

| | |
|---|---|
| FLOW | |
| MINIMUM, MGD | |
| AVERAGE DRY WEATHER, MGD | 21 |
| PEAK WET WEATHER, MGD | 40 |
| $BOD_5$, MG/L | 220 |
| $BOD_5$, 1000 LBS/DAY | 55 |
| SUSPENDED SOLIDS, MG/L | 270 |
| SUSPENDED SOLIDS, 1000 LBS/DAY | 68 |

PILOT PLANT DESIGN DATA

PREAERATION AND GRIT REMOVAL TANKS

| | |
|---|---|
| NUMBER, CONCRETE | |
| NUMBER, CHEMICAL | |
| WIDTH, FT | 14 |
| LENGTH, FT | 55 |
| AVERAGE WATER DEPTH, FT | 11 |
| DETENTION TIME, HR | 1.51 |
| DESIGN FLOW PER TANK, MGD | 2.5 |
| AIR SUPPLIED PER TANK, CU FT/MIN | 6,000 |

LIME FEED EQUIPMENT

| | |
|---|---|
| LIME SLURRY STORAGE TANK | |
| NUMBER | 1 |
| DIAMETER, FT | 12.5 |
| VOLUME, GAL | 8,000 |
| LIME FEED PUMP | |
| NUMBER | 1 |
| CAPACITY, GPM | 75-30 |

CHEMICAL TREATMENT TANKS

| | |
|---|---|
| MIXING CONTROL | |
| NUMBER, CHEMICAL | |
| WIDTH, FT | 7 |
| LENGTH, FT | 7 |
| AVERAGE WATER DEPTH, FT | 7 |
| DETENTION TIME, HRS | 2 |
| OVERFLOW RATE, GAL/SQ FT/DAY | 4,000 |
| DESIGN FLOW PER TANK, MGD | 2.5 |

PRIMARY UNDERFLOW RECIRCULATION PUMP

| | |
|---|---|
| NUMBER | 1 |
| CAPACITY, GPM | 1,500 |

PRIMARY EFFLUENT PUMP

| | |
|---|---|
| NUMBER | 3 |
| CAPACITY, MGD | 4.00 |
| TOTAL HEAD, FT | 15 |
| HORSEPOWER, HP | 75 |

CARBON DIOXIDE FEEDER

| | |
|---|---|
| NUMBER | 1 |
| CAPACITY, LBS/DAY | 1,000 |

CHEMICAL PRIMARY TREATMENT PERFORMANCE

| | |
|---|---|
| $BOD_5$ REMOVAL | |
| PERCENT | 70 |
| LBS/DAY | 1,540 |
| SUSPENDED SOLIDS REMOVAL | |
| PERCENT | 87 |
| LBS/DAY | 1,920 |

PRIMARY EFFLUENT

| | |
|---|---|
| $BOD_5$, LBS/DAY | 660 |
| PRIMARY EFFLUENT TO SECONDARY TREATMENT | |
| FLOW, MGD | 2.2 |
| $BOD_5$, MG/L | 90 |
| SUSPENDED SOLIDS, LBS/DAY | 1,980 |

AERATION TANK

| | |
|---|---|
| NUMBER | 1 |
| WIDTH, FT | 14 |
| LENGTH, FT | 85 |
| AVERAGE WATER DEPTH, FT | 14.33 |
| TOTAL VOLUME, CU FT | 17,000 |
| DESIGN FLOW, GPM | 1,500 |
| DETENTION TIME, HR | 4.5 |
| RETURN SLUDGE RATE, GAL/SQ FT/DAY | 1,000 |
| AIR SUPPLIED PER GAL MIXED LIQUOR | 1.0 |

SECONDARY SEDIMENTATION TANK

| | |
|---|---|
| NUMBER | 1 |
| WIDTH, FT | 14 |
| LENGTH, FT | 85 |
| AVERAGE WATER DEPTH, FT | 10 |
| DESIGN FLOW, GPM | 1,500 |
| OVERFLOW RATE, GAL/SQ FT/DAY | 1,000 |
| AIR SUPPLIED PER GAL EFFLUENT | 0.30 |
| DETENTION TIME, HRS | 4.7 |

RETURN SLUDGE AIR LIFT PUMP

| | |
|---|---|
| NUMBER | 3 |
| CAPACITY, GPM | 750 |
| TOTAL HEAD, FT | 7.5 |

SECONDARY EFFLUENT PUMP

| | |
|---|---|
| NUMBER | 1 |
| CAPACITY, GPM | 400 |
| TOTAL HEAD, FT | 5 |
| HORSEPOWER, HP | 5 |

METHANOL FEED PUMP

| | |
|---|---|
| NUMBER | 1 |
| CAPACITY, GAL/MIN | 2.40 |

DENITRIFICATION TANK

| | |
|---|---|
| NUMBER | 1 |
| DIAMETER, FT | 9.5 |
| HEIGHT, FT | 10.0 |
| AVERAGE WATER DEPTH, FT | 9.3 |
| LIQUID VOLUME, GAL | 8,000 |
| DETENTION TIME, HR | 0.8 |
| DESIGN FLOW, GPM | 160 |

AERATED STABILIZATION TANK

| | |
|---|---|
| NUMBER | 1 |
| WIDTH, FT | 9.8 |
| LENGTH, FT | 10.5 |
| AVERAGE WATER DEPTH, FT | 9 |
| LIQUID VOLUME, GAL | 6,500 |
| DETENTION TIME, HRS | 1.5 |
| DESIGN FLOW, GPM | 150 |

DENITRIFICATION TANK MIXER

| | |
|---|---|
| NUMBER | 1 |
| CAPACITY, RPM | 180 |
| HORSEPOWER, HP | 1 |

FINAL SEDIMENTATION TANK

| | |
|---|---|
| NUMBER | 1 |
| WIDTH, FT | 18 |
| LENGTH, FT | 55 |
| AVERAGE WATER DEPTH, FT | 10 |
| DESIGN FLOW, GPM | 300 |
| OVERFLOW RATE, GAL/SQ FT/DAY | 310 |
| WEIR LOADING RATE, GAL/LIN FT/DAY | 3100 |
| DETENTION TIME, HRS | 6.0 |

RETURN DENITRIFIED SLUDGE AIR LIFT PUMP

| | |
|---|---|
| NUMBER | 1 |
| CAPACITY, GPM | 70-300 |
| TOTAL HEAD, FT | 5.5 |

SECONDARY TREATMENT PERFORMANCE

| | |
|---|---|
| $BOD_5$ REMOVAL | |
| PERCENT | 85 |
| LBS/DAY | 560 |
| SUSPENDED SOLIDS REMOVAL | |
| PERCENT | 80 |
| LBS/DAY | 384 |

OVERALL PILOT PLANT PERFORMANCE

| | |
|---|---|
| $BOD_5$ REMOVAL, PERCENT | 88 |
| SUSPENDED SOLIDS REMOVAL, PERCENT | 98 |
| PILOT PLANT EFFLUENT | |
| $BOD_5$, LBS/DAY | 70 |
| SUSPENDED SOLIDS, LBS/DAY | 48 |

Figure 4-1 ATTF Flow Diagram and Design Data



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## Centrifuges and Related Equipment

Chemical primary sludge was pumped to any one of four 2500-gal. tanks where it was stored until fed to the centrifuge (Fig. 4-2). Carbon dioxide diffusers were installed in two holding tanks to enable pH adjustment of the centrifuge feed. A bar screen removed the rags from the feed sludge, preventing pump clogging or blockage of the centrifuge feed tube. Air diffusers kept the sludge mixed during storage. A similar but much smaller feed and storage system was used in the test work done prior to research grant work.

The usual operating procedure began with operation of the centrifuge in a first stage mode (centrifuging chemical primary sludge) and collection of the centrate in one of the storage tanks. The necessary machine changes were then made, and the machine was run in a second stage mode (centrifuging first stage centrate). Three centrifuges were tested, two extensively.

The Sharples P600 horizontal centrifuge used in the study had a bowl diameter of 6 inches, a bowl length of 14 inches, and a feed rate range of 0 to 12 gpm. The centrifuge feed pump was a Moyno 12-gpm variable speed pump.

The Sharples P3000 horizontal centrifuge used in the study had a bowl diameter of 14 inches, a bowl length of 30 inches, and a feed rate range of 10 to 50 gpm. The centrifuge feed pump was a Moyno 50-gpm variable speed pump.

A Bird solid bowl centrifuge with 6-inch bowl diameter and 12-inch bowl length was employed for limited test work on lime sludges. The feed rate range was 0 to 9 gpm. A Robbins-Meyers Model PS11E feed pump was used.

Polymer was mixed in an 800-gallon tank and pumped to the centrifuge feed tube via a small variable speed chemical pump. Polymer normally was fed internally into the feed zone.

## Pilot Digestion Equipment

The digestion work was carried out on a bench scale, using 5-gal glass bottles for both the reactors and the gas collection vessels. Sludge lines and valves were constructed of polyvinyl chloride, while gas lines were constructed of Tygon tubing. Fig. 4-3 shows a flow schematic of the apparatus. The digesters were mixed by 10-gpm external circulation pumps activated by a timer.

The digesters were placed in a water bath composed of a molded plastic water bed frame kept at a constant temperature by the water bed's thermostatically controlled heating element. Heat input was augmented by a 250-watt

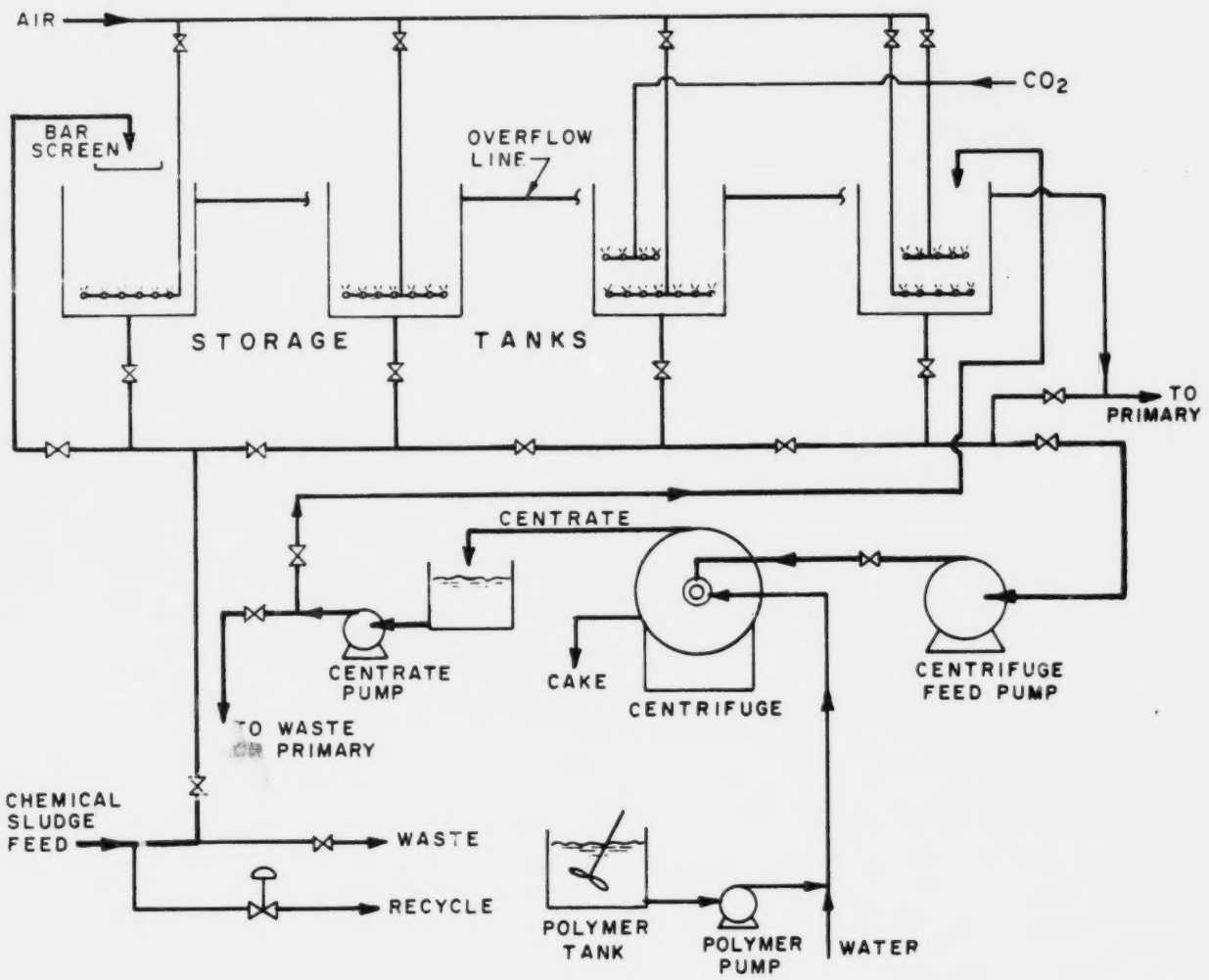


Figure 4-2

Centrifuge Processing Schematic

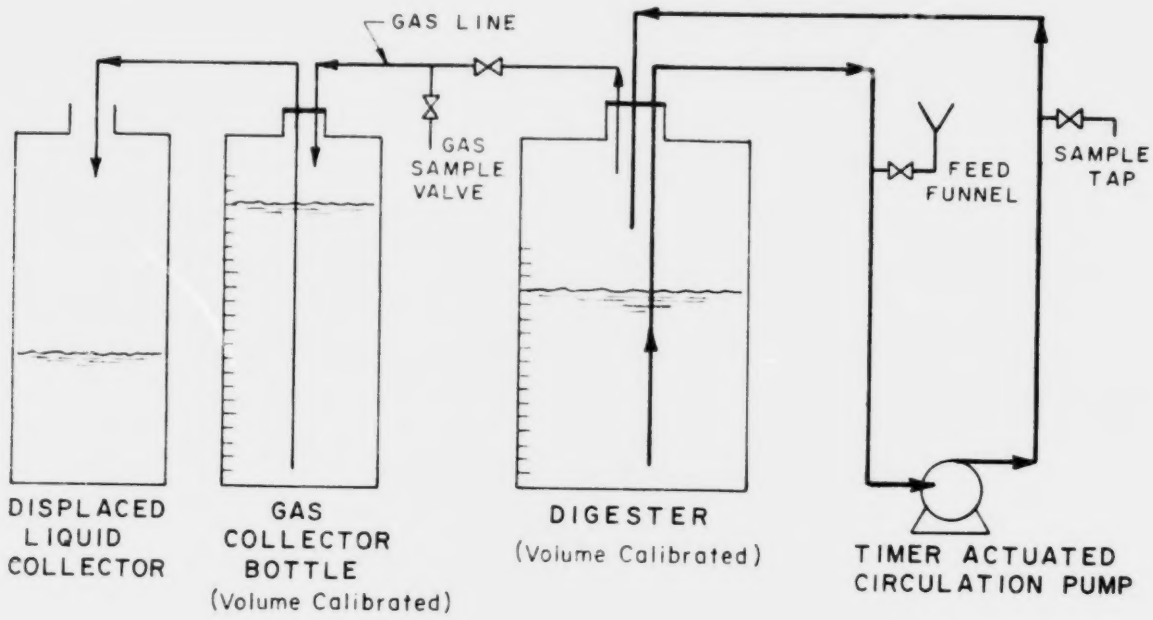


Figure 4-3

Bench Scale Digester Schematic

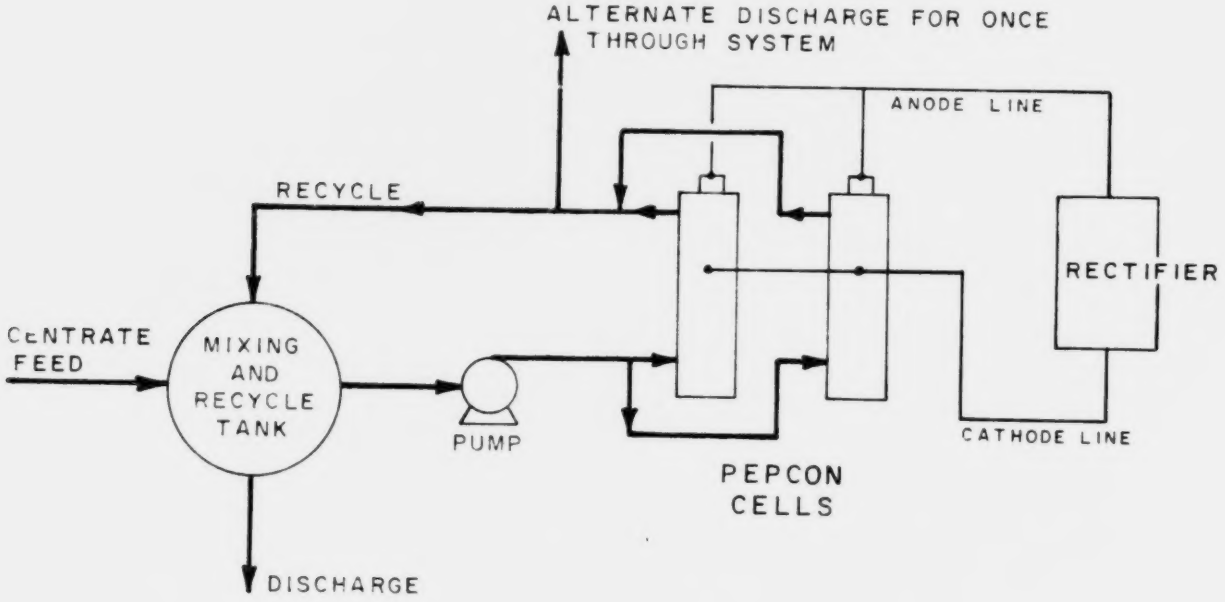


Figure 4-4

Pepcon Treatment Schematic

heating element. The entire water bath was enclosed in a large insulated wooden box, hinged at the top. The heating water was kept well circulated by an external pump. Both water circulation and digester sludge circulation lines were insulated to reduce heat loss. The apparatus maintained sludge temperature within 0.5 C of water bath temperature (normally 38 C).

Electrolytic Test Unit

A Pepcon electrolytic test treatment facility was operated to treat first stage centrate from lime sludges. The unit, produced by Pacific Engineering and Production Nevada, consisted of the following major components:

| | |
|---|---|
| Rectifier | 1000 amp |
| Cells, 4 each, lead dioxide anode | 500 sq in. each<br>2000 sq in. total |
| Cathode, 4 each, copper | |
| Recirculation pumps, 2 each | 24 gpm |
| Recirculation tank | 30 gal |
| Brine tank | 30 gal |

The unit was modified by taking two cells (anode-cathode assembly) out of the flow circuit and connecting the remaining two cells in parallel, both hydraulically and electrically. A schematic of the Pepcon apparatus is shown in Fig. 4-4.

## ANALYTICAL METHODS

A number of analyses on the solids streams were required to determine performance of the various processes as well as component recoveries in the various streams. Most of the analyses were standard techniques as described in Standard Methods for the Examination of Water and Wastewater, 13th Edition. Analyses employed in this project which are listed in Standard Methods are: total and volatile solids; total hardness (EDTA method); calcium hardness (EDTA method); magnesium hardness by difference of total and calcium hardness); soluble phosphate by the stannous chloride method; total calcium, magnesium, iron, and heavy metals by atomic absorption spectrophotometry; total Kjeldahl nitrogen; grease by magnesium sulfate monohydrate dehydration plus Soxhlet extraction with hexane; ammonia nitrogen on centrate of laboratory-centrifuged sludge by the Nesslerization method; and sulfides by the methylene blue visual color-matching method.

Additional analyses performed on sludges were calcium hydroxyapatite, magnesium hydroxide, ferric hydroxide, calcium carbonate, organic matter, and acid insoluble matter. Procedures for these analyses were as follows:

1. A weighed sample of sludge or centrate was dried at 105 C to obtain total solids.

2. A portion of the dried sample was transferred to a crucible, ignited to 1000 C for 2 hours, then collected and weighed to obtain loss on ignition (LOI).

3. A portion of the residue from ignition was dissolved in hydrochloric acid, and the insoluble matter filtered out and weighed to give acid insoluble matter.

4. The filtrate from the above was analyzed for Ca, Mg, Fe (when required) by atomic absorption, and for $PO_4$ by the stannous chloride m. method.

The calculations from the above steps were as follows:

1. From $PO_4$, the calcium hydroxyapatite was calculated. (All $PO_4$ was assumed to be in the hydroxyapatite form).

2. Calcium (net) was determined by deducting from the total the calcium equivalent to calcium hydroxyapatite.

3. The net calcium was calculated to be present as calcium carbonate.

4. Carbon dioxide equivalent to the calcium carbonate was deducted from the LOI matter.

5. Magnesium was calculated as magnesium hydroxide (assuming all Mg present in this form).

6. Water equivalent to magnesium hydroxide was also deducted from the LOI.

7. Where iron was present to a significant degree, iron previously determined was calculated to ferric hydroxide.

8. Water equivalent to ferric hydroxide was also deducted from the LOI.

9. Net LOI was called organic matter.

10. Acid insoluble matter was called the inerts (mostly $SiO_2$).

This analytical method typically yielded recoveries of around 99.9 percent.

Analyses on digested sludge were performed in the following manner:

1. A total solids analysis was made as before.

2. A portion of the dried sample was "wet ashed" with nitric and sulfuric acids, then finally diluted to 250 ml with deionized water.

3. The solution was analyzed for Ca, Mg, and heavy metals by the atomic absorption technique and for $PO_4$ by the stannous chloride method.

4. Total Kjeldahl nitrogen was determined by starting with the original wet sample, using the approach given in Standard Methods.

5. Grease was determined on the original wet sample by the method mentioned earlier.

# SECTION V

# SUMMARY OF LIQUID PROCESSING INVESTIGATION

Interpretation and use of the results of the solids processing investigation requires an understanding of the liquid processing system which generates the solids. A specific concern is the possible operating modes of the liquid processing system.

## ATTF SYSTEM

As previously stated, the ATTF system consists of lime clarification in a chemical primary, followed by nitrification and denitrification. Performance of the primary clarifiers is the main point of concern, since all solids wasted from the process are separated in the primary. Full details on the liquid processing system are available elsewhere[1].

### Chemical Primary

After the raw sewage has been screened, lime is added and flash mixed with recycled primary sludge. When used, ferric chloride and anionic polymer were added at the same point. The sewage is then conveyed to the first element in the treatment system, preaeration. Heavy grit particles, such as sand and unreacted lime, settle out in the preaeration tank. The mild air agitation in the tank just after lime addition promotes the growth of large flocculent particles which settle readily. Operation of the test facility has demonstrated that air mixing in the preaeration tank can be substituted for mechanical mixing and thus avoid fouling problems caused by rags contained in the raw sewage. Rags would tend to wrap around the paddles of mechanical mixers, causing continuous maintenance problems.

Waste activated sludge is also returned to the preaeration tanks for cosettling and concentration in the primary tanks. The ATTF system incorporates no separate waste activated sludge thickening. All of the solids to be processed pass through the primary sedimentation tanks.

Carbon dioxide in the preaeration air has caused a minor operational difficulty; scaling occurs inside the orifices of the coarse bubble diffusers. Initially, the tank had to be drained once every week or two to clean the plugged orifices. The plant staff solved the problem by installing a high pressure air blowoff line and periodically blasted loose the scale. This procedure has extended the period between manual cleaning of the orifices to two months. In the CCSD Water Reclamation Plant preaeration air will be introduced through swing arm diffusers, which will allow maintenance without having to take the tanks out of service.

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After flocculation, the suspension is allowed to settle in rectangular primary tanks. At conventional overflow rates for rectangular primaries (average dry weather surface loading of 1300 gpd/sf and 1.5 hours residence time), the chemical primary far out-performs a parallel conventional primary settling tank (Table 5-1). Peak hourly flows caused the surface loading to reach 1900 gpd/sf.

Primary clarifier operation is managed so that the contractual requirements for industrial reuse on final effluent water quality can be met[1]. To accomplish this, the pH of operation has been varied, and supplemental coagulants utilized to modify effluent qualities. Data for the various modes of operation are shown in Tables 5-2 and 5-3. Of particular interest is the effect of process pH, which determines the required lime dosage. Operation at pH 11.0 in the primary nearly equaled operation at pH 11.5, with one important exception, phosphorus removal (Table 5-2). The mean total phosphorus of 2.3 mg/l in the primary effluent at pH 11.0 caused the final effluent total phosphorus to exceed the 1.0 mg/l limit.

Ferric chloride was tested as a supplemental coagulant to flocculate colloidal phosphorus. At pH 11.0 a 12 mg/l dose of ferric chloride reduced the primary effluent phosphorus to less than 1.0 mg/l (Table 5-3). An even lower lime dose and pH (pH 10.2) is possible if it is compensated by a higher ferric chloride dose (24 mg/l). Under the latter mode of operation, the primary effluent total phosphorus averaged only 0.7 mg/l.

Removals of conventional parameters (BOD, SS, TOC) were all similar for each mode of operation, and therefore in themselves did not dictate a particular method for raw sewage coagulation. However, a substantial difference in the increase of hardness across the primary was observed. For CCCSD sewage, operation at a pH of 11.0 with ferric chloride as a supplemental coagulant yielded the lowest effluent hardness of the coagulation modes tested. While the theoretical minimum solubility point for calcium in clean water is close to pH 10, the level of calcium in the pH 10.2 effluent was as high as at pH 11.0. However, operation at pH 11.0 with ferric chloride produced the greatest removal of magnesium and yielded the least hardness increase across the process.

If hardness is considered a water quality limit, why not use a two-stage recarbonation system, with intermediate settling to enhance hardness removal? Analysis of previous pilot-demonstration work[8] made during the ATTF investigation indicated that little, if any, benefit would be gained from the additional settling and recarbonation stage. During the previous studies the average hardness increase across the two-stage operation varied from -8 to +20 mg/l according to the pH mode. The average for all pH modes was +4. This is very likely due to the effect observed in the present study, namely, that the soluble calcium level is relatively constant with operating

## Table 5-1

## Summary of Primary Sedimentation Performance at Design Flows

| Constituent (mean value) | Raw sewage mg/l | Control primary | | Chemical primary[a] | |
|---|---|---|---|---|---|
| | | mg/l | % removed | mg/l | % removed |
| BOD | 222 | 137 | 38 | 74 | 67 |
| SS | 216 | 64 | 70 | 53 | 75 |
| VSS | - | 57 | - | 31 | - |
| Turbidity | - | 47[b] | - | 32[b] | - |
| TOC | 111 | 77 | 31 | 52 | 53 |
| Soluble organic carbon | 20 | 22 | -10 | 29 | -45 |
| Total phosphorus as P | 9.7 | - | - | 2.8 | 71 |
| Orthophosphate as P | 8.5 | - | - | 1.8 | 79 |
| Settleable solids | 9[c] | 0.8[c] | 91 | 0.12[c] | 99 |
| Calcium hardness[e] | 68 | - | - | 153 | - |
| Magnesium hardness[e] | 113 | - | - | 66 | - |
| Grease | 55[d] | 39 | 29 | 16 | 71 |

[a] Operation from February 29 to March 9 with average flow of 2.66 mgd at pH 11.0 without ferric chloride.

[b] JTU

[c] ml/l

[d] Not measured, typical value shown.

[e] as $CaCO_3$

## Table 5-2

## Summary of Limed Primary Sedimentation Performance

| Constituent (mean value) | pH 11.5 operation[a] $Ca(OH)_2$: 500 mg/l | | | | | pH 11.0 operation[b] $Ca(OH)_2$: 400 mg/l | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| | Raw sewage mg/l | Control primary | | Chemical primary | | Raw sewage mg/l | Control primary | | Chemical primary | |
| | | mg/l | % removed | mg/l | % removed | | mg/l | % removed | mg/l | % removed |
| $BOD_5$ | 190 | 103 | 46 | 50 | 74 | 192 | 121 | 37 | 60 | 69 |
| SS | 199 | 57 | 71 | 41 | 79 | 195 | 57 | 71 | 47 | 76 |
| VSS | - | 43 | - | 20 | - | - | 46 | - | 25 | - |
| Turbidity | - | 35[c] | - | 16[c] | - | - | 40[c] | - | 26[c] | - |
| TOC | 107 | 59 | 45 | 37 | 65 | 118 | 68 | 42 | 48 | 59 |
| Soluble organic carbon | 16 | 16 | 0 | 23.4 | -46 | 17 | 21 | -24 | 28 | -65 |
| Total phosphorus as P | 9.4 | - | - | 0.96 | 90 | 9.2 | - | - | 2.3 | 75 |
| Settleable solids | 8.2[d] | 0.3[d] | 95 | nil | 100 | 9.5[d] | 0.2[d] | 98[d] | <0.1[d] | >99 |
| Calcium hardness[e] | 76 | - | - | 168 | - | 76 | - | - | 156 | - |
| Magnesium hardness[e] | 96 | - | - | 40 | - | 103 | - | - | 59 | - |
| Hardness increase[e] | - | - | - | 36 | -28 | - | - | - | 36 | -20 |
| Grease | 57 | 32 | 44 | 12 | 79 | 53 | 42 | 21 | 19 | 64 |

[a] December 22, 1971 to February 10, 1972, at average flow 1.30 mgd.

[b] February 11 to February 28, 1972, at average flow 1.12 mgd.

[c] JTU

[d] ml/l

[e] as $CaCO_3$

# Table 5-3

## Summary of Lime/Ferric Primary Sedimentation Performance

| Constituent (mean value) | pH 11.0 operation[a] $Ca(OH)_2$: 400; $FeCl_3$: 14 | | | | | pH 10.2 operation[b] $Ca(OH)_2$: 289; $FeCl_3$: 24 | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| | Raw sewage mg/l | Control primary | | Chemical primary | | Raw sewage mg/l | Control primary | | Chemical primary | |
| | | mg/l | % removed | mg/l | % removed | | mg/l | % removed | mg/l | % removed |
| $BOD_5$ | 210 | 109 | 48 | 53 | 75 | 178 | 106 | 40 | 59 | 66 |
| SS | 305 | 69 | 78 | 27 | 91 | 235 | 59 | 75 | 31 | 87 |
| VSS | - | 49 | - | 14 | - | - | 48 | | 19 | - |
| Turbidity | - | 45[c] | - | 14[c] | - | - | 41[c] | - | 15[c] | - |
| TOC | 130 | 72 | 45 | 37 | 72 | 117 | 68 | 42 | 43 | 63 |
| Soluble organic carbon | 23 | 18 | 21 | 24 | -4 | 20 | 17 | 15 | 25 | -25 |
| Total phosphorus as P | 9.5 | - | - | 0.85 | 91 | 9.4 | - | - | 0.68 | 93 |
| Settleable solids | 13.2[d] | 0.13[d] | 99 | nil | 100 | 12.1[d] | .1 | 99 | <0.1[d] | >99 |
| Calcium hardness as $CaCO_3$ | 75.5 | - | - | 139 | - | 72 | - | - | 148 | - |
| Magnesium hardness as $CaCO_3$ | 90 | - | - | 32.5 | - | 90 | - | - | 70 | - |
| Hardness increase as $CaCO_3$ | - | - | - | 6 | -4 | - | - | - | 56 | -34 |
| Grease | 146 | - | - | 9.5 | 94 | 66 | - | - | [illegible] | [illegible] |

[a] March 23, 1972 to April 5, 1972 at average flow 1.19 mgd.

[b] April 7, 1972 to April 30, 1972 at average flow 1.20 mgd.

[c] JTU

[d] ml/l

pH at CCCSD. It also follows that little calcium carbonate would be available for recalcination from a second stage precipitator at CCCSD.

All of the previously reported data were for periods when underflow solids were recycled to the preaeration basin. The objective of recycling solids is improvement of the rate of the precipitation reactions involving calcium, magnesium and phosphorus. Lime reactor TSS was maintained in the range of 1000 to 2500 mg/l. To test the effect of solids recycle on the primary overflow, solids recycle was discontinued from June 3 to June 22, 1972. The major performance change observed was in effluent hardness. During this period of operation at pH 11.0 with a ferric chloride dose of 10 to 16 mg/l, the hardness increase across the primary averaged 32 mg/l. This compares to the hardness increase of 6 mg/l for March 23 through April 5, when underflow solids were recycled to hold flocculator solids in the range of 900 to 1500 mg/l TSS (Table 5-3). When flocculator solids were increased to 1700 to 3900 mg/l for the period of April 25 to 30, the hardness increase across the primary was 16 mg/l. On the basis of hardness data obtained to date, it appears desirable at CCCSD to recycle underflow solids at a rate which will maintain lime reactor TSS in the range of 900 to 1500 mg/l when operating at pH 11.0.

Nitrification

After primary clarification, the chemical primary effluent is passed directly to the oxidation-nitrification tanks with no intervening recarbonation stage. The chief source of carbon dioxide ($CO_2$) for recarbonation is the $CO_2$ generated in the nitrification process. Carbon dioxide is derived from the oxidation of both carbon and (through an intermediate step) nitrogen[1]. This biologically-produced $CO_2$ is usually sufficient for recarbonation without the need for addition of $CO_2$ from external sources. When needed, external carbon dioxide is vaporized from liquid storage and fed directly to the first bay of the oxidation-nitrification tank.

Primary clarifier performance is controlled to maximize organic removal in the primary, thereby controlling sludge age in the oxidation-nitrification tank. In this way, the longer sludge ages (greater than 10 days) required for year-round nitrification can be attained without separating the carbonaceous oxidation and nitrification stages as is done in the three-sludge system[9,10].

Lime clarification ahead of nitrification also provides a significant buffer against heavy metals that would be toxic to the nitrifying organisms. High-lime treatment is one of the most efficient processes for heavy metals removal[8,23].

Denitrification

A denitrification system was developed at the ATTF which couples an anoxic residence period for nitrate reduction with an aerobic period for liquid and

sludge stabilization[1]. This process proved to be so efficient for nitrogen removal that the effluent contains on the average less than 2 mg/l of total nitrogen, compared to approximately 30 mg/l in the raw sewage to the plant.

Results

Performance data for a representative three months of operation are shown in Table 5-4. Of particular interest is the fact that the 90 percentile performance level for the various constituents does not vary widely from the median performance level. This provides statistical confirmation of the ATTF system's operating stability.

Table 5-4 shows that the concentration of organics, as measured by BOD and organic carbon, is low in both the nitrified and denitrified effluents. Operation of the activated sludge process for complete nitrification results in high organic removals. Suspended matter in the nitrified and denitrified effluents is also exceptionally low. This is attributed in part to the fact that the sludge has had a high sludge volume index (Table 5-5). Effluents from sludges with a high sludge volume index (SVI) are quite commonly very clear. For instance, Keeter, et al[17] found that effluent solids decreased linearly with SVI up to the point of clarifier failure. The mechanism for this enhanced clarification is flocculation of dispersed particles in contact with the sludge as it settles.

Nutrients are effectively removed in the ATTF system. Total nitrogen in the denitrified effluent averages less than 2 mg/l. More than half of the effluent nitrogen is organic nitrogen, and a major fraction of that will be removed by effluent filtration. Total phosphorus is below the 1 mg/l contractual limit.

Also of interest are the changes in alkalinity which occur through the process (Table 5-4). Lime treatment causes a net increase in alkalinity of the raw sewage by adding hydroxide ion (even though removing phosphorus alkalinity). Nitrification, which is an acid producing process, causes a decrease in alkalinity. Denitrification is a base (hydroxide ion) producing process, which causes another increase in alkalinity. The net effect of all these processes is a decrease in the total alkalinity of the water through the process.

# Table 5-4

## ATTF Performance Summary, April 16 - July 15, 1972

| Constituent | Raw sewage, mg/l | | | Chemical primary,[b] mg/l | | | Nitrified effluent, mg/l | | | Denitrified effluent, mg/l | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | mean | median | 90%[a] | mean | median | 90%[a] | mean | median | 90%[a] | mean | median | 90%[a] |
| $BOD_5$ | 203 | 199 | 235 | 57 | 54 | 79 | 3.6 | 3.5 | 6.8 | 3.2 | 3.0 | 4.8 |
| TOC | 122 | 120 | 152 | 42 | 41 | 55 | 8.9 | 8.5 | 11.5 | 9.5 | 9.0 | 11 |
| SOC | 22 | 23 | 25 | 27 | 27 | 31 | 5.6 | 5.5 | 6.7 | 5.9 | 6.0 | 6.6 |
| SS | 214 | 212 | 295 | 26 | 23 | 45 | 4.5 | 4.0 | 7.8 | 4.5 | 4.0 | 7.7 |
| Turbidity | - | - | - | 12.8[c] | 12.0[c] | 23[c] | 1.4[c] | 1.3[c] | 2.2[c] | 1.4[c] | 1.3[c] | 1.8[c] |
| Settleable solids | 13.5[d] | 13.0[d] | 16.5[d] | .084[d] | 0[d] | .37[d] | - | - | - | - | - | - |
| Organic N[e] | - | - | - | - | - | - | .26 | 0 | .67 | 1.1 | 1.1 | 2.5 |
| $NH_3$ - N | - | - | - | 24.0 | 23.8 | 28.5 | .48 | .30 | .58 | .31 | .30 | .40 |
| $NO_2$ - N | - | - | - | - | - | - | 27 | 27 | 32 | .48 | 0 | .79 |
| $NO_2$ - N | - | - | - | - | - | - | .015 | .013 | .027 | .009 | .008 | .018 |
| Total P | 9.86 | 9.57 | 11.19 | .86 | .59 | 1.85 | 1.04[f] | .72[f] | 2.11[f] | .50 | .36 | .98 |
| Ortho P | 9.74 | 9.90 | 11.24 | .61 | .46 | 1.75 | 1.00 | .71 | 1.80 | .52 | .36 | .90 |
| TDS | 583 | 561 | 750 | - | - | - | 634 | 636 | 724 | 537 | 551 | 616 |
| Conductivity | 1230[g] | 1210[g] | 1366[g] | - | - | - | 1226[g] | 1223[g] | 1394[g] | 1160[g] | 1147[g] | 1318[g] |
| Alkalinity[h,i] | 215 | 218 | 237 | 254 | 249 | 293 | 105 | 106 | 127 | 183 | 187 | 217 |
| Ca[h] | 67.9 | 66.0 | 78.5 | 155 | 150 | 184 | 159[k] | 161[k] | 187[k] | 174[l] | 172[l] | 199[l] |
| Mg[h] | 84.0 | 84.0 | 95.6 | 38.6 | 34.0 | 74.0 | 57.3[k] | 61.0[k] | 73.0[k] | 30.1 | 30.0[l] | 48.4[l] |
| Temp range | 16.5-23.0[j] | - | - | - | - | - | - | - | - | 16-24[l] | - | - |

[a] 90% of observations are equal to or less than stated value.

[b] pH 10.2 operation to June 1, pH 11.0 thereafter.

[c] Jackson Turbidity Units

[d] ml/l

[e] June 27 to July 15 only.

[f] April 16 to June 1 only.

[g] Micromhos

[h] as $CaCO_3$

[i] April 16 to May 31 only.

[j] degrees C.

[k] April 16 to May 31 only.

[l] June 1 to July 15 only.

## Table 5-5

## Process Parameters for Nitrification and Denitrification

| Parameter | Nitrification | Denitrification |
|---|---|---|
| Flow, mgd | .47 ± .04 | .47 ± .04 |
| Residence time, hr | 6.5 (aeration) | .8 2 (reactor)<br>.79 (stabilization) |
| MLVSS, mg/l | 2985 ± 388 | 3115 ± 449 |
| VSS | 83.8 | 83.6 |
| SVI, ml/g | 294 ± 57 | 250 ± 50 |
| SRT, day | 21.3 ± 5.2 | 6.5 ± 23.6[a] |
| N oxid or rem lb/lb MLVSS/day | .038 ± .008 | .13 ± .05 |
| TOC rem lb/lb MLVSS/ day | .047 ± .010 | - |
| $BOD_5$ and NOD load lb/ 1000 cu ft/day | 38 | - |
| M/N ratio | | 3.54 ± 1.07 |

[a]May 1 to June 16 only, on April 16 to 30 and June 17 to July 15 wasting from activated sludge into the denitrification system.

## SECTION VI
## SLUDGE PRODUCTION AND THICKENING

There are two interfaces between the ATTF liquid processing and sludge processing systems: one is at the point where waste solids stream enters the sludge processing sequence, the other is at the point where residual solids return to the main flow stream (such as in a centrate). The problem of residual solids is discussed in later sections. The present section deals with the quantity and quality of sludge entering the processing sequence.

### SLUDGE PRODUCTION BY PRECIPITATION

All sludge processed in the ATTF system is cycled through the primary and appears in the primary underflow. The sources of sludge include the raw sewage suspended solids, the solids that are wasted to the chemical primary from the subsequent biological treatment stages, and, of course, the inorganic chemical sludges that are precipitated due to lime and iron addition. The latter are composed of calcium carbonate, magnesium hydroxide, ferric hydroxide, calcium phosphate compounds, and a host of minor constituents such as precipitates of heavy metals.

#### Measured Sludge Production

Sludge production was monitored in the ATTF system by setting the waste sludge pumping rate at a known (and recorded) rate and by monitoring sludge total solids by means of three grab samples per day. Sludge production is tabulated in Table 6-1 for four different coagulation modes tested in the chemical primary.

#### Calculated Sludge Production

Sludge production can also be calculated from analyses of influent and effluent samples using the data on suspended solids, phosphorus, magnesium, and calcium, combined with chemical feed data. This was done as follows:

1. Suspended solids (SS) removed in the primary was taken as the difference between the raw influent and effluent values (sampled 5 days a week).

2. Waste activated sludge (WAS) removed in the primary was calculated from the metered quantities of sludge returned to the chemical primary. During most of the test periods shown, activated sludge was wasted to the denitrification system (to build up biological solids) rather than to the primary.

3. Phosphorus removed in the process was assumed to be precipitated as hydroxyapatite ($CA_5$ (OH) $(PO_4)_3$).

4. Magnesium removed in the process was assumed to be in the form of magnesium hydroxide ($Mg(OH)_2$). Magnesium was sampled five days a week.

5. Due to the high pH of coagulation, iron, when added to the process as ferric chloride ($FeCl_3$), was presumed to precipitate as ferric hydroxide ($Fe(OH)_3$) rather than as a phosphate compound.

6. Calcium carbonate precipitation was calculated based on a mass balance around the calcium ion. The equation used was:

$$CaCO_3 \text{ precip} = Ca_{in} + Ca_{dose} - Ca_{out} - Cap;$$

where each component is expressed as calcium carbonate.

Data on calcium in the influent ($Ca_{in}$) and in the effluent ($Ca_{out}$) were available from weekday analysis. Lime dose was measured daily. The lime requirement for phosphorus was calculated from the mass of phosphorus removed, assuming that the phosphorus was precipitated as hydroxyapatite.

The results of these balances are presented in Table 6-1 along with the ratios of measured sludge production to calculated production. The maximum difference between measured and calculated production rates was 15 percent for the four modes of operation. Further, two modes had less measured production than calculated, while in two modes measured production exceeded calculated production. This would indicate that there is no systematic error in the method of calculation of the mass of sludge production.

One significant source of error affecting these comparisons is that of measuring actual sludge production. Only three grab samples were taken per day, and accuracy in the rate of sludge pumping was based on the frequency of operator attention, which was variable.

Comparisons between the periods are interesting; however, it should be noted that raw suspended solids varied significantly through the test periods and affected sludge quantities. Comparisons in sludge quantities for a common set of of raw sewage characteristics are made in Section XI.

Since the calculated and measured quantities are in substantial agreement the engineer can construct with confidence a solids balance for a lime precipitation system using pilot plant data on the constituents. Carefully conducted jar tests should also yield adequate data provided that the need for a

## Table 6-1

## Summary of Sludge Production for Several Modes of Chemical Primary Operation

| Parameter \ Operating pH, Chemical dose[a] | pH 11.5[b] $Ca(OH)_2$:500 | pH 11.0[c] $Ca(OH)_2$:400 | pH 11.0[d] $Ca(OH)_2$:400 $FeCl_3$:11 | pH 10.2[e] $Ca(OH)_2$:289 $FeCl_3$:24 |
|---|---|---|---|---|
| Sludge production | | | | |
| lb per mil gal | 6930 | 6190 | 6125 | 3875 |
| mg/l | 832 | 743 | 735 | 465 |
| Calculated sludge production, mg/l | | | | |
| SS | 182 | 171 | 278 | 204 |
| WAS | 0 | 7 | 0 | 3 |
| $Ca_5(OH)(PO_4)_3$ | 42 | 37 | 47 | 47 |
| $Mg(OH)_2$ | 34 | 31 | 33 | 12 |
| $Fe(OH)_3$ | 0 | 0 | 9 | 16 |
| $CaCO_3$ | 541 | 419 | 431 | 268 |
| Total | 799 | 665 | 798 | 550 |
| Ratio measured production to calculated production | 1.04 | 1.12 | 0.92 | 0.85 |

[a] Dose in mg/l.

[b] Dec 23, 1971 to Feb 11, 1972.

[c] Feb 13 to March 21, 1972

[d] March 23 to April 5, 1972.

[e] April 7 to April 30, 1972.

reasonable scale-up in chemical dose is recognized. The authors have used both pilot data and jar test results to construct the solids balance for a number of proposed treatment plants.

## Sludge Composition

The calculated distribution between sludge components shown in Table 6-1 can be compared to sludge analyses of grab samples collected during the evaluation periods (Table 6-2). The sludge analyses were made in connection with the centrifuge test work reported in later sections. The samples were normally taken between 9 am and 5 pm, with most in the afternoon.

Before the comparisons could be made, the computed compositions in Table 6-1 had to be modified to take into account several factors not considered in the computation. First, while very little activated sludge was wasted to the primary during any of the periods shown in Table 6-1, activated sludge was wasted to the primary on those individual days when centrifuge tests were made. This wasting was in approximate proportion to the levels that will be maintained in the full-scale plant. As a consequence, approximately 60 mg/l of waste activated sludge was in the sludge sampled and not in the constituent balances. Second, in the constitutent balances it was assumed that all of the effluent SS were organic in nature. While no specific tests were conducted on the effluent SS, other data (Table 5-4) indicates that about 20 mg/l of $CaCO_3$ was lost over the weirs. Lastly, in the sludge analyses, organics were taken as volatile matter corrected for ignition losses related to calcium carbonate and magnesium hydroxide. To make comparisons to the sludge analyses, SS in the constituent balance were assumed to be 80 percent volatile. Once these factors were included in the constituent balance, the fractions could be computed and compared to the grab sample sludge analyses.

Considering the fact that the sludge samples were grab samples while the constituent balances were based on composites, there is reasonable agreement between sludge analysis and computations. The grab samples would reflect transient diurnal fluctuations and daytime conditions which would normally be expected to differ somewhat from a 24-hour composite.

Ferric hydroxide was consistently less in the analyses than in calculated compositions. It is believed that is due to the method of ferric hydroxide feed; the chemical was fed at a constant rate despite the fact that chemical primary flows fluctuated diurnally. Grab samples in the afternoon, therefore, would have lower iron content than the calculated compositions, as the latter would reflect 24 hours of sludge generation.

## Biological Sludge Generation

Biological sludge production was monitored in the ATTI system to provide a basis for projecting sludge generation in the full-scale water reclamation plant.

## Table 6-2

## Comparison of Calculated Sludge Composition to Grab Sample Analysis

| Mode of primary operation and date | Constituent, percent | | | | | |
|---|---|---|---|---|---|---|
| | Organic | $Ca_5(OH)(PO_4)_3$ | $Mg(OH)_2$ | $Fe(OH)_3$ | $CaCO_3$ | Balance |
| pH 11.5, no $FeCl_3$ | | | | | | |
| Feb 1, 1972 | 25.7 | 5.8 | 6.7 | 0 | 55.3 | 6.8 |
| Feb 1 | 26.0 | 5.5 | 6.0 | 0 | 55.0 | 6.9 |
| Feb 1 | 29.2 | 5.7 | 7.6 | 0 | 51.5 | 5.9 |
| Feb 1 | 28.4 | 5.9 | 6.1 | 0 | 53.3 | 6.4 |
| Feb 2 | 34.0 | 5.3 | 7.9 | 0 | 46.5 | 6.3 |
| Feb 7 | 30.0 | 7.0 | 6.0 | 0 | 51.0 | 6.0 |
| Mean | 28.9 | 5.9 | 6.7 | 0 | 52.1 | 6.4 |
| (Calculated) | (24.4) | ( 4.8) | (3.9) | ( 0) | (60.6) | ( 6.0) |
| pH 11.0, no $FeCl_3$ | | | | | | |
| Feb 14, 1972 | 33.0 | 8.0 | 4.0 | 0 | 49.0 | 5.0 |
| (Calculated) | (28.0) | ( 5.1) | (4.3) | ( 0) | (55.5) | ( 6.9) |
| pH 11.0 with $FeCl_3$ | | | | | | |
| April 5, 1972 | 31.0 | 6.4 | 4.5 | 0.4 | 47.2 | 10.5 |
| (Calculated) | (33.3) | ( 5.4) | (3.8) | (1.0) | (47.9) | ( 8.3) |
| pH 10.2 with $FeCl_3$ | | | | | | |
| April 24, 1972 | 36.6 | 10.2 | 1.4 | 0.8 | 40.0 | 11.0 |
| April 24 | 35.1 | 10.1 | 1.6 | 2.3 | 45.4 | 5.5 |
| April 27 | 32.8 | 10.8 | 1.8 | 1.0 | 48.7 | 4.9 |
| April 27 | 32.0 | 12.2 | 1.7 | 1.6 | 49.7 | 2.8 |
| Mean | 34.3 | 10.8 | 1.7 | 1.3 | 46.0 | 6.0 |
| (Calculated) | (37.4) | ( 7.7) | (1.9) | (2.6) | (40.8) | ( 9.3) |

## Table 6-3

## Nitrification Sludge Production and Yield

| Month, 1972 | lb BOD removed per day | lb SS produced per day[a] | Yield, lbs SS produced per lb BOD removed |
|---|---|---|---|
| January | 188 | 89 | 0.47 |
| February | 158 | 106 | 0.67 |
| March | 266 | 139 | 0.52 |
| April | 211 | 156 | 0.74 |
| May | 243 | 181 | 0.74 |
| June | 163 | 150 | 0.92 |
| Mean | 205 | 139 | 0.67 |

[a]Includes solids wasted, solids lost over weirs, and changes in sludge inventory.

Sludge production during the food processing season (July to October) is not presented herein because it represented atypical conditions. The cannery ceased operation in 1972 and no longer will be a contributor to the CCCSD system.

Nitrification Sludge. The net sludge yield in the oxidation-nitrification stage of the system for six months of operation averaged 0.67 lb SS per lb BOD removed (Table 6-3). Waste activated sludge production for the full-scale 30-mgd plant is projected in Section XI.

Denitrification Sludge. As has been described in detail elsewhere[1], the denitrification system was initially operated in the conventional manner (Fig. 6-1), but operation in this mode did not produce an acceptable effluent. The system was modified to incorporate an aerated stabilization tank between the reactor and clarifier. As has been reported, this reduced sludge yields to the point where the only wasting was a 4 mg/l loss over the effluent weirs, compared to 14 mg/l obtained for the unmodified system or 16 mg/l predicted by sludge yield data for other denitrification systems[1].

The data previously reported[1] was reanalyzed to develop sludge yield data (Table 6-4). The conventional system had a sludge yield of approximately 0.6 lb SS per lb nitrate-N. This is within the range of 0.2 to 0.9 obtained by others[12] for suspended growth denitrification. After modification, the sludge yield dropped to 0.07, well below that for the conventional system or the value of 0.4 suggested by Christensen and Harremoes[12] in their comprehensive literature review. Thus, it can be seen that the modified denitrification system produced only 12 to 17 percent of the sludge produced in conventional denitrification systems. The result of this is that the denitrification system has essentially no impact on solids processing in the ATTF system.

## SLUDGE THICKENING IN THE CHEMICAL PRIMARY

The chemical primary has two basic functions: the first is separation of the solids from the liquid, and the second is thickening of these solids. While the ATTF solids processing sequence incorporates separate thickening stages (Fig. 3-2), considerable thickening takes place in the primary tank itself.

### General Observations

Thickening depends primarily on solids loading on the clarifier bottom, which can be expressed as lbs per sq ft per day. The solids loading depends on the rate of flow and solids content. In turn, solids content depends on the amount of solids precipitated plus the amount purposely recycled from the underflow.

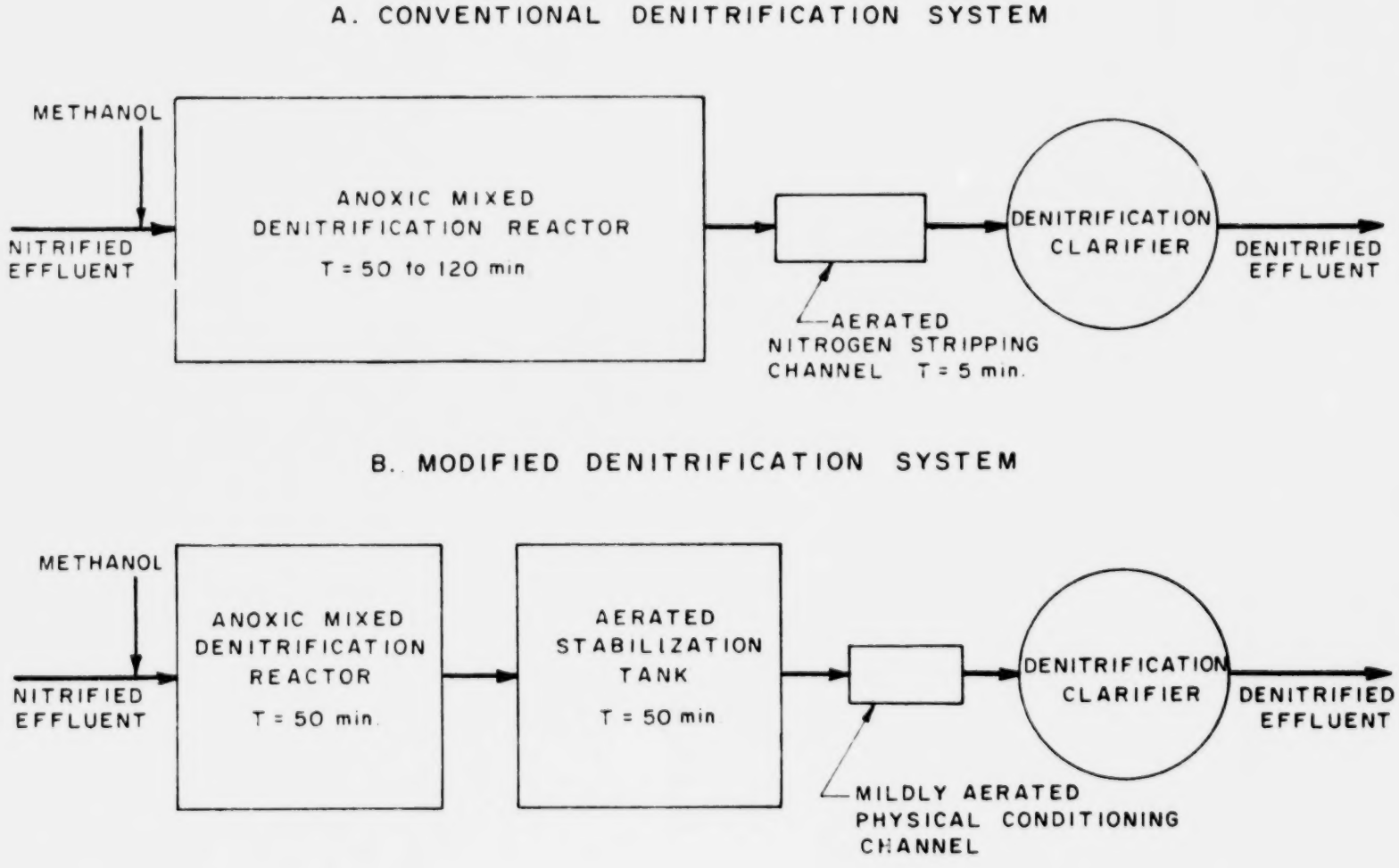



Figure 6-1

Comparisons of Denitrification Systems

## Table 6-4

## Comparison of Sludge Yields for Conventional and Modified Denitrification Systems

| Denitrification system and test period | Nitrate-N removed lb per day | SS wasted, lb per day | | | Yield, lb SS per lb Nitrate-N |
|---|---|---|---|---|---|
| | | Effluent | Change in inventory | Total | |
| Conventional system[b] March 3 to 13, 1972 | 106 | 60 | a | 60 | .6 |
| Modified system[b] May 2 to June 16, 1972 | 104 | 15.8 | -8.3 | 7.5 | .07 |

[a] Inventory record indicates an increase in inventory by as much as 20 lbs/day. Record is somewhat erratic and therefore inventory change is neglected.

[b] Refer to Fig. 6-1.

## Table 6-5

## Summary of Chemical Primary Operating Conditions During Thickening Test Periods

| Test period | Primary operating pH | Waste activated sludge | Supplemental ferric chloride addition | Supplemental polymer addition |
|---|---|---|---|---|
| Dec 1, 1971 to Feb 10, 1972 | 11.5 | no | no | no |
| Feb 11 to Mar 23, 1972 | 11.0 | yes | no | no |
| March 25 to April 5, 1972 | 11.0 | no | yes | no |
| April 7 to April 26, 1972 | 10.2 | yes | no | no |
| April 26 to May 8, 1972 | 10.2 | yes | yes | no |
| May 9 to May 21, 1972 | 10.2 | yes | yes | yes |
| May 25 to May 31, 1972 | 10.2 | yes | yes | yes |
| June 24 to July 15, 1972 | 11.0 | yes | yes | no |
| July 15 to Oct 15, 1972 | 11.0 | yes | yes | no |

Solids content was routinely monitored by means of composited grab samples taken from the reactor (preaeration tank). Solids loading could then be calculated using the reactor solids level and flow data. Fig. 6-2 shows the relationship between flow and reactor solids level. This figure proved useful in interpreting the thickening data.

It was found that the inventory of sludge in the primary sedimentation tank varied diurnally. The amount of sludge in the tank was at a minimum early in the morning, increasing to a maximum in the afternoon. Fig. 6-3 shows two typical sludge profiles. The change in inventory is due to the fact that sludge production varies diurnally, while the rate of sludge pumping was held constant. On some occasions there would be no sludge at sampling station No. 2 in the morning, resulting in very thin sludge in the underflow.

From the surface of the tank it could be seen that the sludge settled out very rapidly. The sludge profiles confirmed that the bulk of the sludge settled out in the first three-eighths of the primary tank. Sludge was seldom found past Station 5 unless thickening failure had occurred.

Whenever sludge intruded to Station 6 or beyond, under the effluent weirs, large flocs immediately began appearing in the effluent. As long as the sludge layer was held in the first half of the tank no large flocs appeared. This factor helps explain the observed stability of these clarifiers as a function of flow rate.[1]

Expansion of the sludge layer into the effluent end of the primary was taken as evidence of thickening failure, since no significant increase in underflow solids concentration occurred at the time of the layer expansion. On this basis, it is estimated that only 50 percent of the tank floor is effective in thickening. Since the calculated solids loadings (Fig. 6-2) always included the total floor area, actual solids loadings are at least twice those calculated.

Observation of sludge thickening in the primary was complicated by several factors. These were:

1. Limiting solids loadings are difficult to assess. To establish actual limiting loadings, a failure in the thickening must occur. Up to the point of thickening failure, the TS attained in the underflow is determined by the rate of pumping, i.e., by strictly a dilution relationship. Therefore, unless thickening failure occurs, daily measurements of underflow solids and associated computed solids loadings do not necessarily provide useful information about thickening capability.

2. Because sludge was pumped at a constant rate, composite underflow samples included thin night and early morning samples which diluted the heavier afternoon and evening samples.

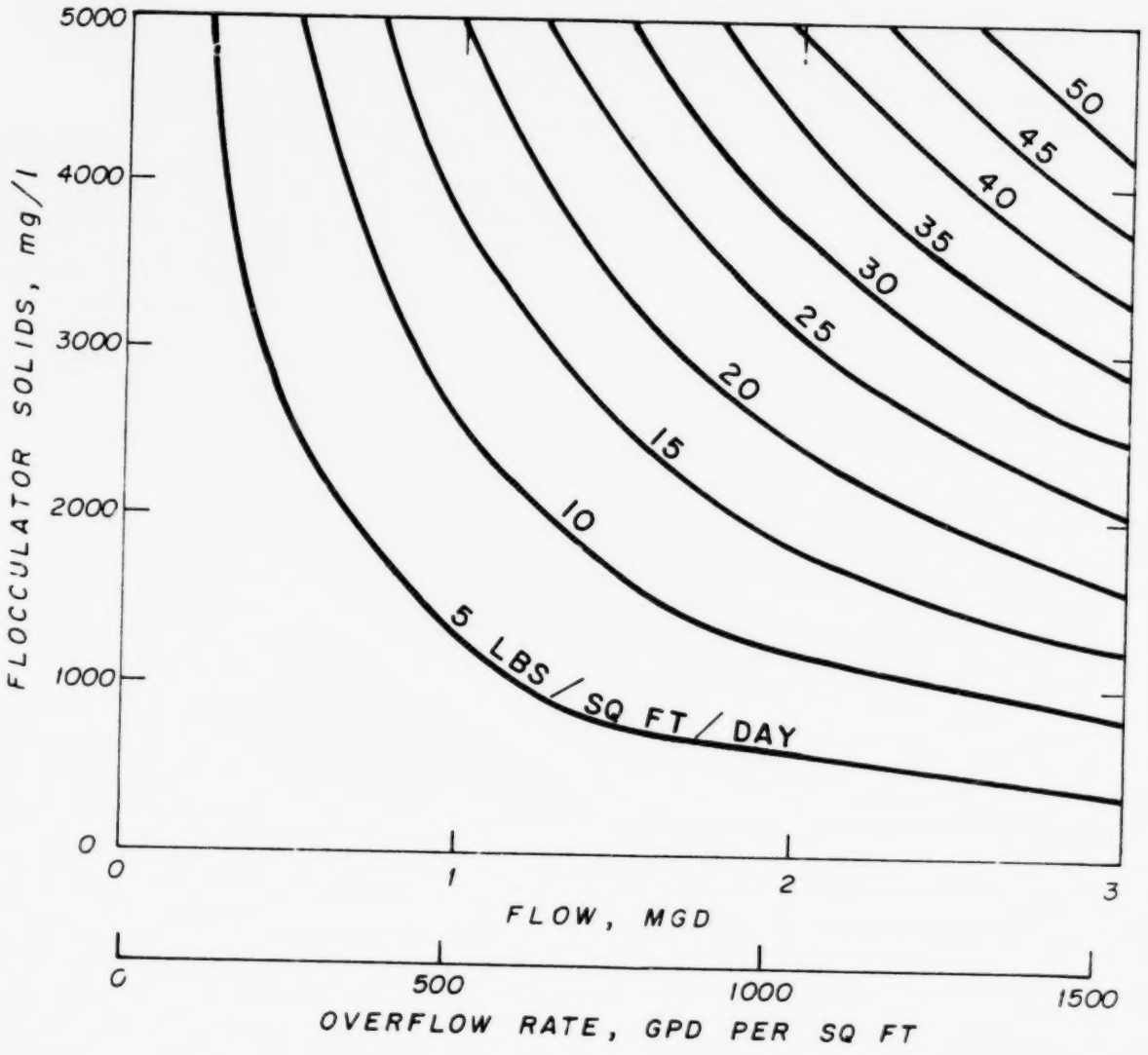



Figure 6-2

Clarifier Solids Loading as a Function of Flow and Flocculator Solids Level

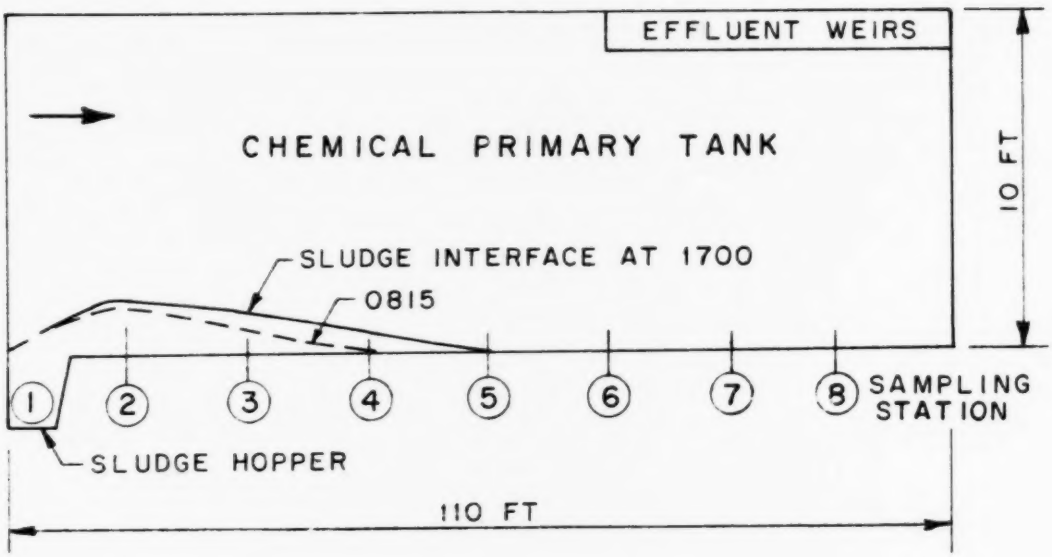



Figure 6-3

Sludge Profile on October 24, 1972

3. Operation of the primary was dictated by an experimental program which was not always compatible with the study of sludge thickening.

Thickening data for the primary clarifier are summarized according to operating period in the following sections. Table 6-5 summarizes operating conditions in the chemical primary for these periods.

December 1, 1971 to February 10, 1972, pH 11.5

During this period no thickening failures were observed, and limiting solids loadings could therefore not be definitely established. Maximum attained underflow concentrations are listed:

| Solids loading lb per sq ft per day | Underflow TS, percent |
|---|---|
| 10 | 8.2 |
| 7 | 9 |

At about 7 percent TS and above, coning and bridging occurred in the sludge hoppers, and septic sludge zones developed. With steeper hoppers and better agitation this should be less of a problem.

February 1 to March 23, 1972, pH 11.0

There was no thickening failure during this period. Maximum attained underflow concentrations are listed:

| Solids loading lb per sq ft per day | Underflow TS, percent |
|---|---|
| 18 | 6.6 |
| 16 | 8.0 |
| 7 | 9.9 |

March 23 to April 5, 1972, pH 11.0

Sludge profile monitoring during this period indicated that the sludge was pumped thin. No failure in thickening occurred. At 8 lbs per sq ft per day solids loading, up to 4.8 percent TS was attained.

April 7 to April 26, 1972, pH 10.2

Thickening failure occurred when the underflow waste rate was adjusted to a rate inconsistent with the level to which the sludge could be thickened. The limiting total solids level appeared to be 4.2 percent. This sludge could be thickened to 4.2 percent up to at least a solids loading of 14 lb per sq ft per day.

April 26 to May 8, 1972, pH 10.2

No thickening failure was observed during this period. At 12 lb per sq ft per day, the underflow solids could be thickened up to 3.9 percent TS.

May 9 to May 21, 1972, pH 10.2

Use of a polymer, ICI Atlasep 3A3 at a dose of 0.25 mg/l, yielded improvements in sludge thickening over previous operation at pH 10.2. At a solids loading of 16 lb per sq ft per day, up to 6.0 percent TS was attained. No thickening failure was observed.

May 25 to May 31, 1972, pH 10.2

Use of an alternate polymer, Dow A-23 at 0.29 mg/l, allowed solids loadings to be increased to 29 lb per sq ft per day at underflow TS of 4.3 percent. While the Dow A-23 was not evaluated at the same solids loadings as the Atlasep 3A3, it did not appear to be as effective.

June 1 to June 23, 1972

During this period underflow was purposely pumped thin and no significant observations on thickening could be made.

June 24 to July 15, 1972, pH 11.0

Sludge during this period could not be thickened as well as in previous comparable periods (e.g., Feb 11 to Mar 23 or Mar 23 to April 15, 1972). Based on observations of sludge profiles at thickening failure, these limiting levels are estimated:

| Solids loadings lb per sq ft per day | Underflow TS, percent |
|---|---|
| 20 | 3 |
| 14 | 4.3 |

The main difference between this period and the similar previous periods was sewage temperature. During this period sewage temperature was between 20 and 24 C, while in the two previous periods the temperature was usually between 15 and 18 C. It is possible that the chemical sludges do not thicken as well at higher temperatures. This effect may be due to increased biological activity in the sludge (see next section).

July 15 to October 15, 1972, pH 11.0

During this period a cannery processing peaches and tomatos was contributing to the system. Thickening in the primary during this period was quite erratic.

Biological action formed gas bubbles in the solids very quickly, which tended to make the solids less compact. Underflow solids were usually in the range of 2.5 to 4.5 percent TS. Sewage temperatures ranged from 20 to 24 C.

Under these conditions sludge could be thickened to 4.0 percent at 20 lb per sq ft per day. There were a few days when sludge could be thickened to 5.0 percent at even higher loadings.

Summary

In general, sludges with a high pH (11.0 and 11.5) thickened better than the sludge generated at pH 10.2. However, thickening of the latter sludge could be significantly improved by the use of a small dose (0.25 ppm) of anionic polymer. During the summer, under elevated temperatures, biological activity caused "gassing" of the sludge and thickening deteriorated.

## SECTION VII

## WET CLASSIFICATION OF COMBINED SLUDGES

Most of the lime used for raw sewage coagulation in the ATTF system is precipitated as calcium carbonate. The amount of new lime required in the ATTF system can be reduced by recovering, recalcining, and reusing the precipitated calcium carbonate. Recalcining can be readily accomplished in a multiple hearth furnace.

The problem in recovering lime from raw sewage sludge is that many constituents are precipitated with the sludge besides calcium carbonate. As explained in Section VI, organics, phosphorus, magnesium, iron, and other constituents are coprecipitated with calcium carbonate. These constituents, if left with the calcium carbonate during recalcining, would be returned to the process. Eventually, these "inerts" would build up to such a magnitude that the solids processing facilities would be overwhelmed. This problem has been the major reason that lime recovery in the primary stage of application has not been practiced [18]. The problem of lime recovery has also been a deterrent to the use of lime in raw sewage coagulation.

The system developed for lime recovery employs a centrifuge for separation of the calcium carbonate from the other sludge constituents. This process is termed "wet classification." The fundamental mechanism for classification is based on the fact that the calcium carbonate settles more rapidly than any of the other sludge constituents. By purposely operating the centrifuge at incomplete solids capture, the calcium carbonate can be preferentially recovered in the cake while the other solids are preferentially rejected to the centrate. The investigation of wet classification was concerned with the operating variables to optimize the separation of calcium carbonate from the other solids.

A second aspect of the investigation which is integrally related to classification was the dewatering of the separated calcium carbonate. Efforts were directed toward reducing the water content of the calcium carbonate-rich sludge to minimize the energy requirements for recalcination.

Solid bowl centrifuges manufactured by Sharples division of Pennwalt Corporation were used for studying wet classification. Two machines were used, the P600 and P3000. The P600 has a maximum capacity of approximately 10 gpm of feed, while the P3000 can process a maximum of about 50 gpm of feed.

Runs were made under various operating conditions to obtain data for designing the solids handling system for the CCCSD Water Reclamation plant. The experi-

Rental design was oriented towards flexibility of operation so that several variables could be changed in the experiments at once. Thus the modes tested were based on an ongoing evaluation of the test results and on the investigator's broad experience in centrifuge operation.

Under each mode of operation the centrifuge was operated sufficiently long to assure a steady state condition. Since the residence time in a centrifuge is short, a steady state condition when operating to classify the sludge is attained in five to ten minutes. After reaching a steady state condition, the primary sludge (centrifuge feed), the water-rich effluent (centrate) and the solids concentrate (cake) were sampled. The analytical methods used for analyzing the samples are described in detail in Section IV.

The recovery of total solids was calculated by material balance using the following equation:

$$R_t = \frac{c_t}{f_t} \frac{(f_t - e_t)}{c_t - e_t} 100$$

where:

$R_t$ = recovery of total solids, percent

$c_t$ = total solids in cake, percent

$e_t$ = total solids in centrate, percent

$f_t$ = total solids in feed, percent

The term $\frac{f_t - e_t}{c_t - e_t}$ gives the fraction of total feed stream that is separated in the cake stream. Multiplying the concentration of solids in the cake by the fraction separated as cake gives the amount of solids recovered in the cake stream for a unit of feed. Dividing the amount of solids in the cake by the amount of solids in a unit of feed gives the fractional recovery of solids. Recovery of total solids was reported as percent.

After calculating the total solids recovery, the recovery of an individual constituent can be calculated by the following equation:

$$R_c = \frac{c_c R_t}{f_c}$$

where:

$R_c$ = recovery of a given constituent, percent

$R_t$ = recovery of total solids, percent

$c_c$ = constituent in total solids of cake, percent

$f_c$ = constituent in total solids of feed, percent

## CONSTITUENT SEPARATION

The purpose of wet classification is to maximize recovery of calcium carbonate while rejecting the other components to as great an extent as possible. The factors affecting the classification of the solid components are discussed below. A subsequent section of the report discusses dewatering of the calcium carbonate-rich cake.

### Separation Efficiency

The calcium carbonate crystals settle more rapidly than the other solids. Consequently, a partial recovery of the total solids with a centrifuge concentrates the calcium carbonate in the cake. To get an indication of the efficiency of component separation, the recovery of each component was compared with the total solids recovery. The plots are shown in Fig. 7-1 for all of the runs listed in Table 7-1.

There is a considerable scatter of data resulting from different degrees of separation and from experimental error. However, trend lines can be drawn through the data points when it is recognized that two boundary conditions must be satisfied. These are:

1. At 100 percent recovery of total solids all constituents must be completely removed.

2. At 0 percent recovery of total solids, no individual constituent can be recovered.

These recovery curves are summarized in Fig. 7-2. From the figure, it can be observed that there is a fairly broad range in which fairly high calcium carbonate recovery is coupled with reasonable classification. For instance, at 50 percent total solids recovery, 70 percent of the calcium carbonate is recovered with less than 30 percent of the other major constituents, while at 70 percent total solids recovery, almost 90 percent of the calcium carbonate is recovered with less than 50 percent of the other major constituents.

### Centrifuge Variables

Besides demonstrating the feasibility of wet classification, the data presented in Fig. 7-1 show the wide range in degree of separation attained. The variations in separation are due primarily to differences in centrifuge operating conditions. The principal variables are feed rate, centrifugal force, and pond setting.

# Table 7-1

## Run Data for Wet Classification

| Process Parameters | | | | | | | | Recoveries in Cake, percent | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Run | Machine | Centrifugal force per lb mass, g's | Auger speed, ΔRPM | Auger pitch, in. | Pond number | Feed rate, gpm | Flocculation pH | Total solids | Calcium carbonate | Magnesium hydroxide | Hydroxy-apatite | Ferric hydroxide | Organic matter | Other materials |
| 2-7 | Sharples P-600 | 2100 | 50 | 2 | 1 | 8 | 11.5 | 67 | 90 | 30 | 23 | | 41 | 57 |
| 2-5 | | | | | | 10 | | 63 | 86 | 27 | 20 | | 41 | 51 |
| 2-1 | | | | | 3 | 10 | | 69 | 90 | 43 | 25 | | 43 | 57 |
| 2-4 | | | | | | 12 | | 65 | 78 | 35 | 24 | | 49 | 87 |
| 2-32 | | 1050 | 35 | | 1 | 9 | | 66 | 84 | 66 | 18 | | 59 | 44 |
| 2-12[a] | | 2100 | 16 | | 3 | 3 | | 86 | ~ 100 | 51 | 76 | | 65 | 60 |
| 8-1 | Sharples P-3000 | 2100 | 26 | 3 | 1 | 23 | 11.0 | 64 | 96 | 44 | 10 | | 50 | 66 |
| 8-4 | | | | | | 49 | | 56 | 89 | 34 | 11 | | 42 | 60 |
| 8-12 | | | | | 2 | 25 | | 69 | 88 | 64 | 30 | | 43 | 67 |
| 8-9 | | | | | | 49 | | 47 | 57 | 31 | 13 | | 32 | 50 |
| 8-8 | | | | | 3 | 21 | | 72 | 96 | 58 | 34 | | 56 | 66 |
| 8-5 | | | | | | 49 | | 64 | 75 | 45 | 41 | | 55 | 66 |
| 8-41 | | 1500 | | | 1 | 19 | | 61 | 85 | 36 | 11 | | 42 | 76 |
| 8-40 | | 1500 | | | | 43 | | 55 | 76 | 32 | 9 | | 35 | 74 |
| 4-3-72 | Sharples P-600 | 3050 | 65 | 1 | 1 | 10 | | 62 | 87 | 30 | 23 | 31 | 38 | 56 |
| 4-134 | | 2100 | 50 | | | 5 | 10.2 | 57 | 79 | 38 | 23 | 23 | 41 | 41 |
| 4-106 | | | | | | 7 | | 48 | 69 | 21 | 18 | 6 | 33 | 52 |
| 4-105 | | | | | | 11 | | 49 | 77 | 32 | 16 | 12 | 35 | 30 |
| 4-123[b] | | | 9 | | 1 | 4 | | 97 | ~ 100 | 97 | 89 | 87 | 91 | ~ 100 |

[a] Three pounds of Allied Chemical colloid 726 polymer added per ton of dry solids in the feed.

[b] Two and a half pounds of ICI America Atlasep 2A2 polymer added per ton of dry solids in feed.

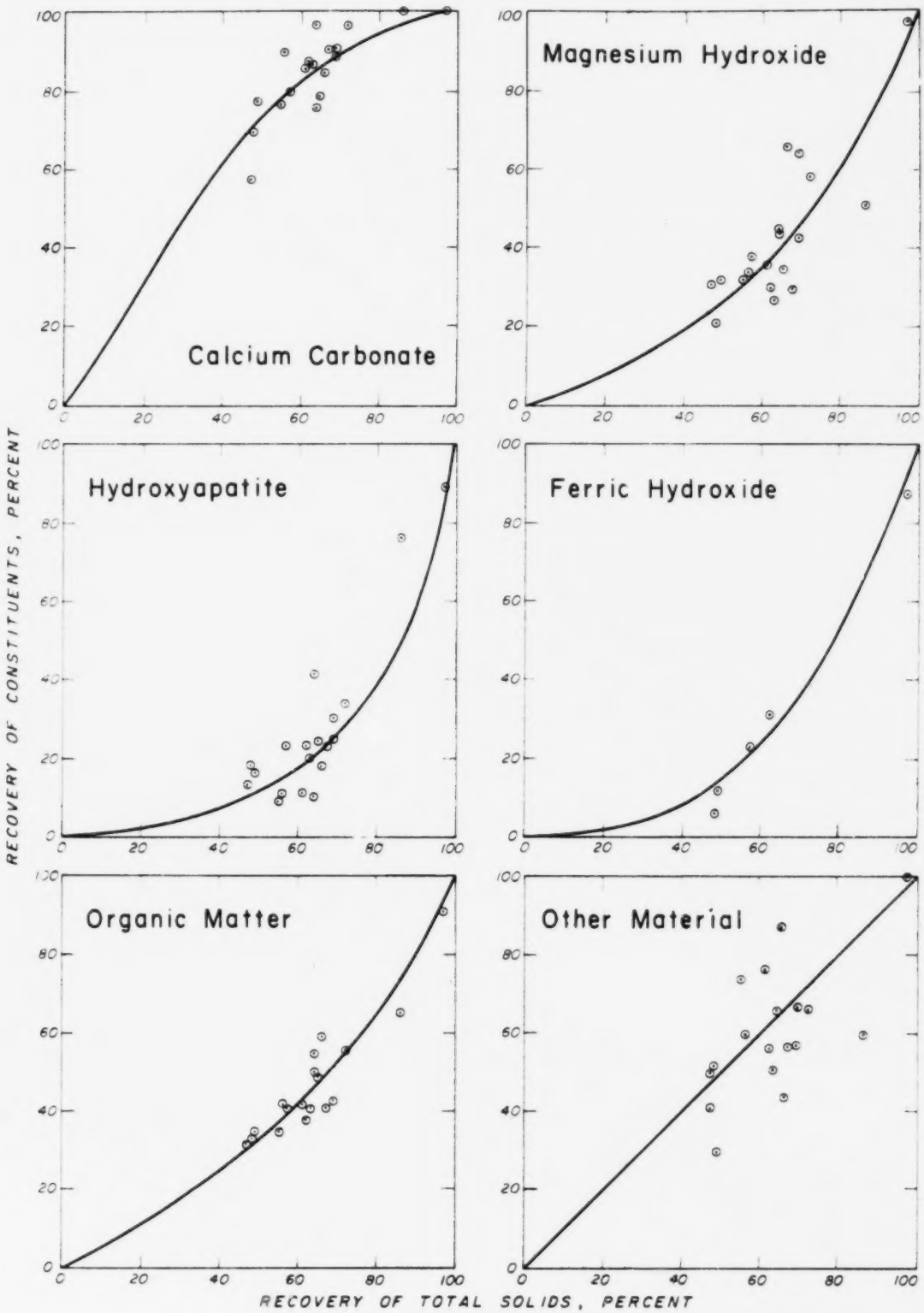



Figure 7-1 Constituent Recovery During Classification

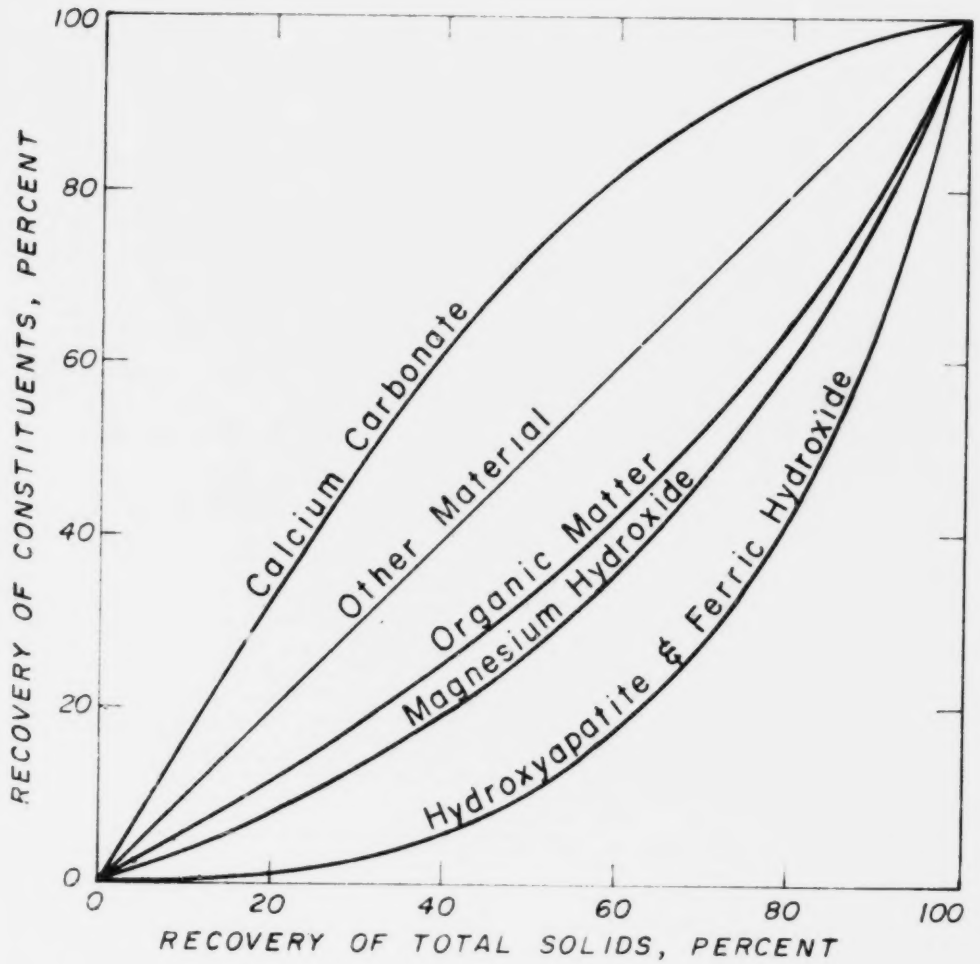



Figure 7-2

Summary of Constituent Recoveries During Wet Classification

Feed Rate. One of the most important parameters in the [illegible] is the feed rate and the effect of variations in feed rate. Fortunately, the variation in feed rate could be tolerated for the purpose of constituent separations. Examination of the data in Table 7-1 reveals that for runs made under the same conditions, except for feed rate, there was only a slight reduction in constituent separation with an increase in feed rate.

Centrifugal Force. Another important variable in centrifuge operation is the centrifugal force. Based on experience at Stanford Research Institute, most of the runs were made at a centrifugal force per unit mass of 1200 times the acceleration of gravity (g). However, a few runs were made at different g-levels. Over the range of about 1000 to 3000 g, recovery of calcium carbonate increased only slightly with an increase in centrifugal force. But newsprint and the high calcium carbonate clay, as will be discussed in a subsequent [illegible] section, centrifugal force is an important operating parameter.

Pond Setting. The depth of the liquid layer in the centrifuge bowl would be expected to be an important variable. Increasing the depth of the liquid (lower pond number) definitely had an adverse effect on the separation of constituents. The best constituent separation was always attained at the deepest pond setting. For example, runs 8-4 and 8-5 (Table 7-1) were all process parameters constant except pond setting. The pond setting of [illegible] resulted in a much higher recovery of calcium carbonate and a greater rejection of all other components.

Lime Dosage, or pH

There is some indication that calcium carbonate is more easily separated from the other components at a high pH than at a low pH. However, other variables were also changed simultaneously with pH, and the pH effect could therefore not be firmly established. Runs made at pH of 10.2 and 11.5 were made with different machine speeds, while runs made at pH 11.0 were made with the larger of the two machines. Also, the concentration of the various constituents in the sludge was substantially different at the various pH levels (refer to Section XI), and that also could account for the differences in degree of separation.

Polymer Addition

Two runs were made with polymer addition to increase the capture of solids. Because of the limited amount of data, it is difficult to determine the effect of polymer on classification. In run 2-12, the recovery of calcium carbonate and rejection of the other components were better than would be expected, based on the relationship presented in Fig. 7-2 between total solids recovery and recovery of individual constituents. For the other run (4-123) a 97 percent recovery of total solids was obtained, and consequently, the recovery of all constituents was high, resulting in little classification.

Based on results from run 2-12, in which an 86 percent recovery of total solids was obtained, it appears that a polymer can be used in wet classification to increase both the total solids and calcium carbonate recoveries.

## DEWATERING AND RECOVERING SOLIDS

Besides separating the individual constituents of the solids, the dewatering of the recovered cake must also be considered. Factors which are important are the centrifuge operating conditions and the lime dosage or pH level used in the primary treatment from which the sludge is produced.

### Centrifuge Variables

As with the separation of solid constituents, the extent of dewatering is affected by several centrifuge operating variables. The effects of feed rate, centrifugal force, and pond setting were studied to determine their relative importance.

Feed Rate. As was noted in the discussion of constituent separation, feed rate can be varied over a wide range with only a minor effect on separation. Feed rate was also found to have little effect on total solids recovery at all pH levels (Figs. 7-3, 7-4, and 7-5). For the smaller of the two machines tested, the P600, a change in feed rate from 4 to 12 gpm, a threefold change, resulted in less than a 10 percent decrease in total solids recovery. Approximately the same effect of feed rate was observed with the larger machine, the P3000. The cake dryness seemed to improve slightly with an increase in feed rate, but the improvement was too small to be considered significant.

Centrifugal Force. For constituent separations centrifugal force seemed to have a slight influence. For dewatering and solids recovery, however, the tests showed that centrifugal force is an important variable. Increasing the centrifugal force improved dewatering and increased solids recovery (Figs. 7-3, 7-4, 7-6, and 7-7). A centrifugal force of 2100g appears to be adequate for dewatering.

Pond Setting. The liquid depth (pond setting) had a minor effect on total solids recovery but a substantial effect on cake dryness (Fig. 7-6). For dewatering, as was previously observed for constituent separation, the best results were obtained with as shallow a liquid depth as possible (pond setting No. 1).

Comparison of Machines. A few runs were made with both the P600 and P3000 during the same time period. The results are presented in Figs. 7-7 and 7-8. The observed difference in performance of the two machines is possibly due to differences in augers and auger speeds rather than machine size. The auger

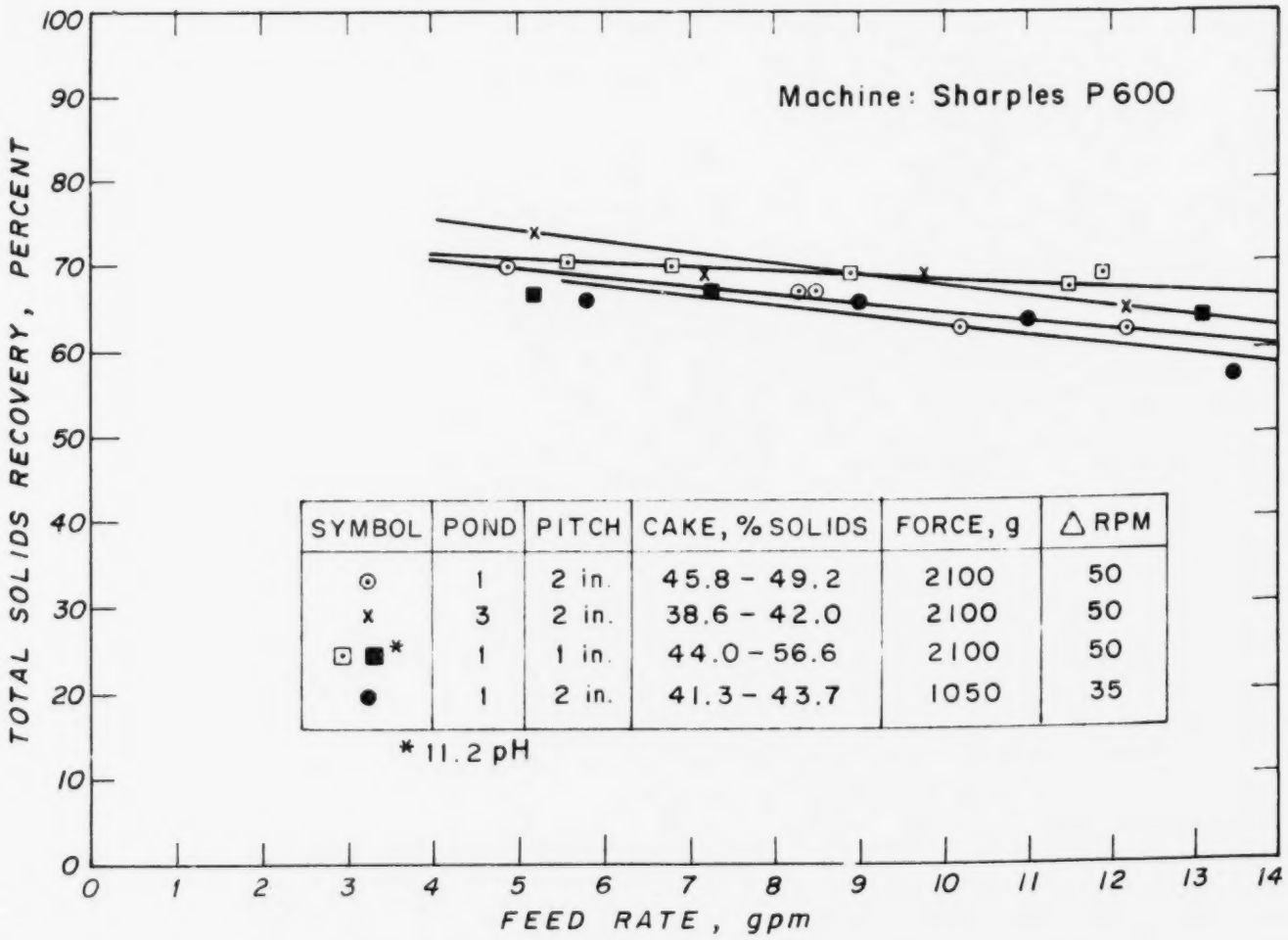



Figure 7-3 Solids Recovery at Flocculation pH of 11.5

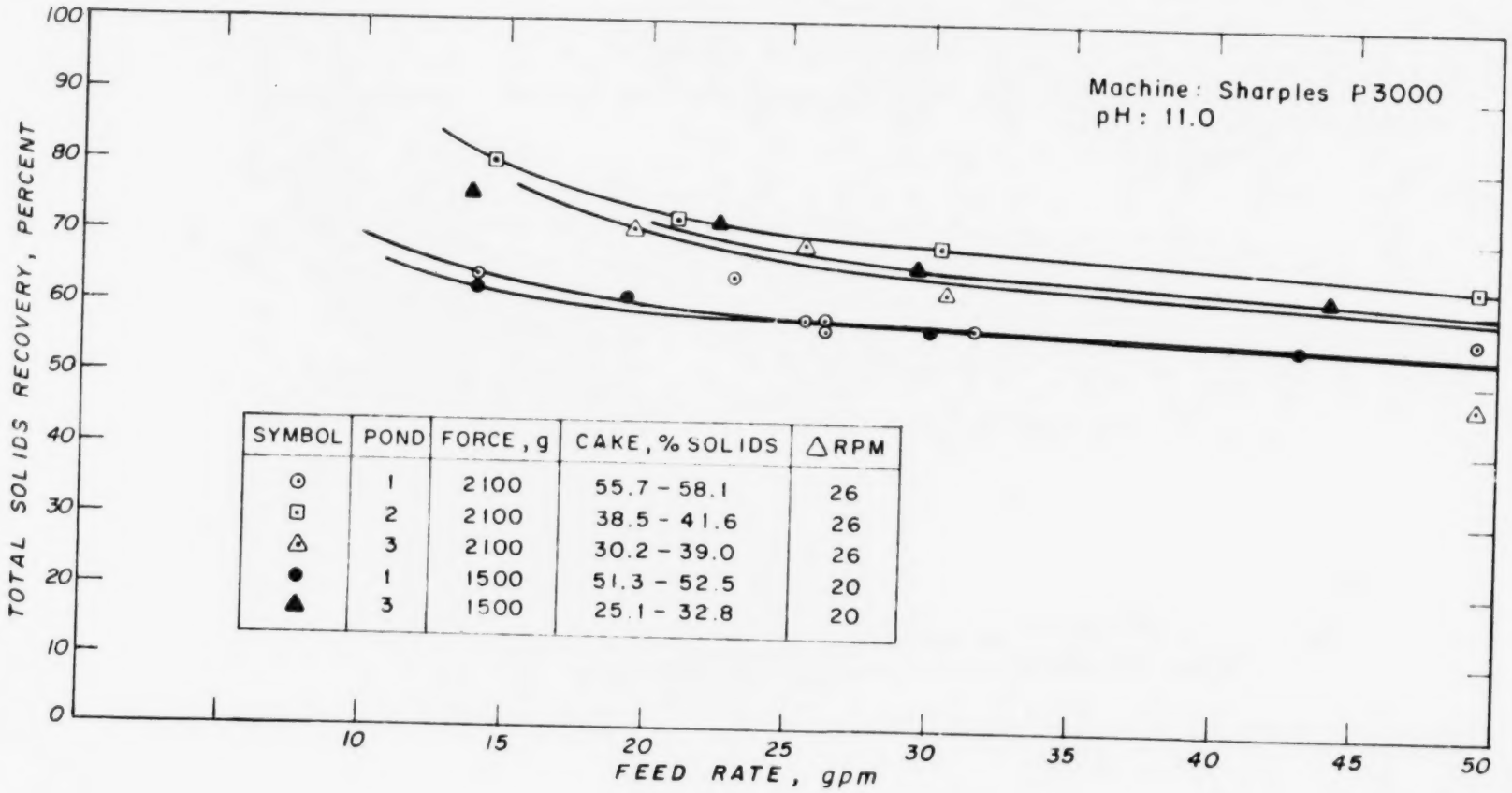



Figure 7-4

Solids Recovery at Flocculation pH of 11.0

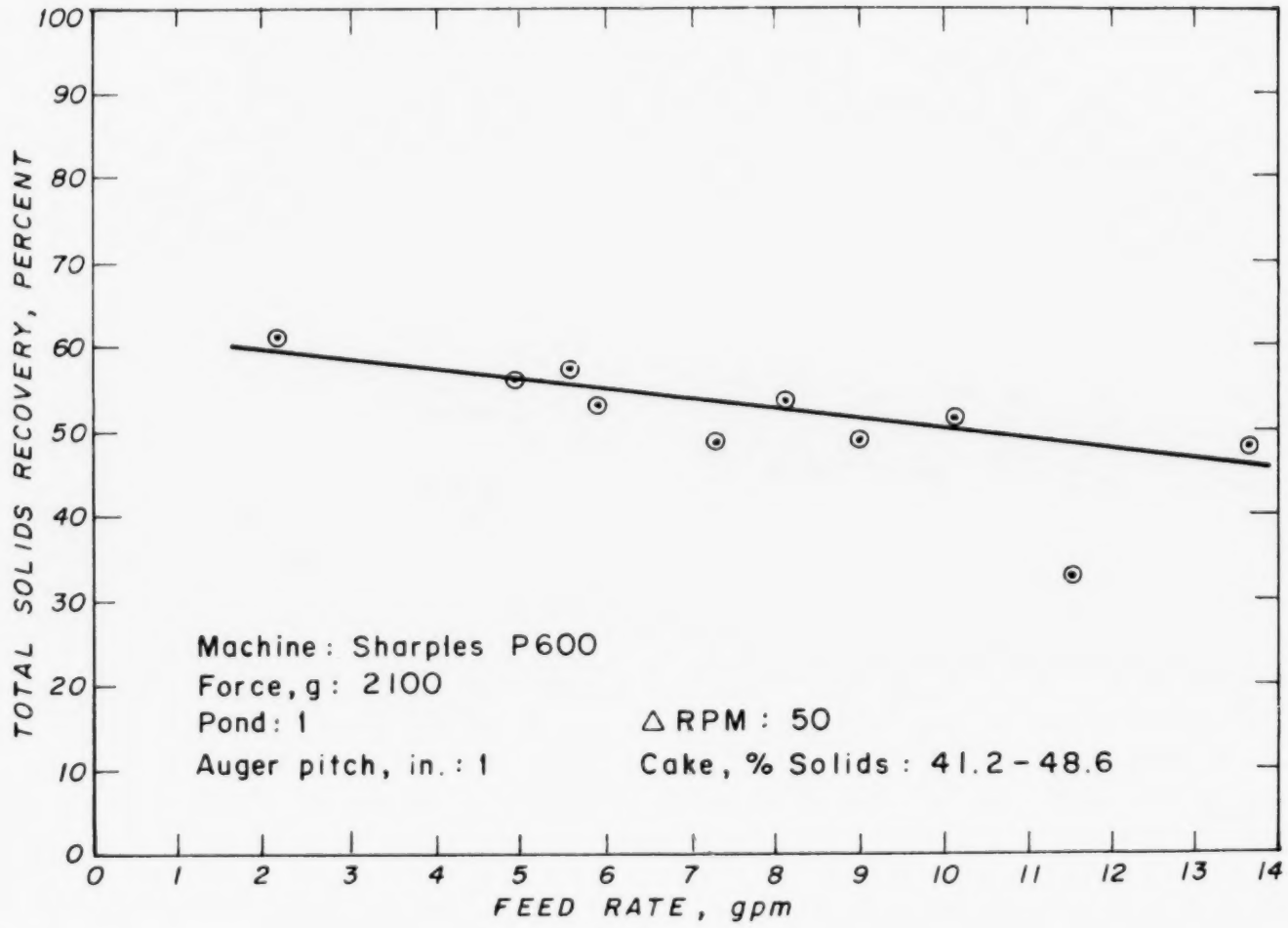



Figure 7-5

Solids Recovery at Flocculation pH of 10.2

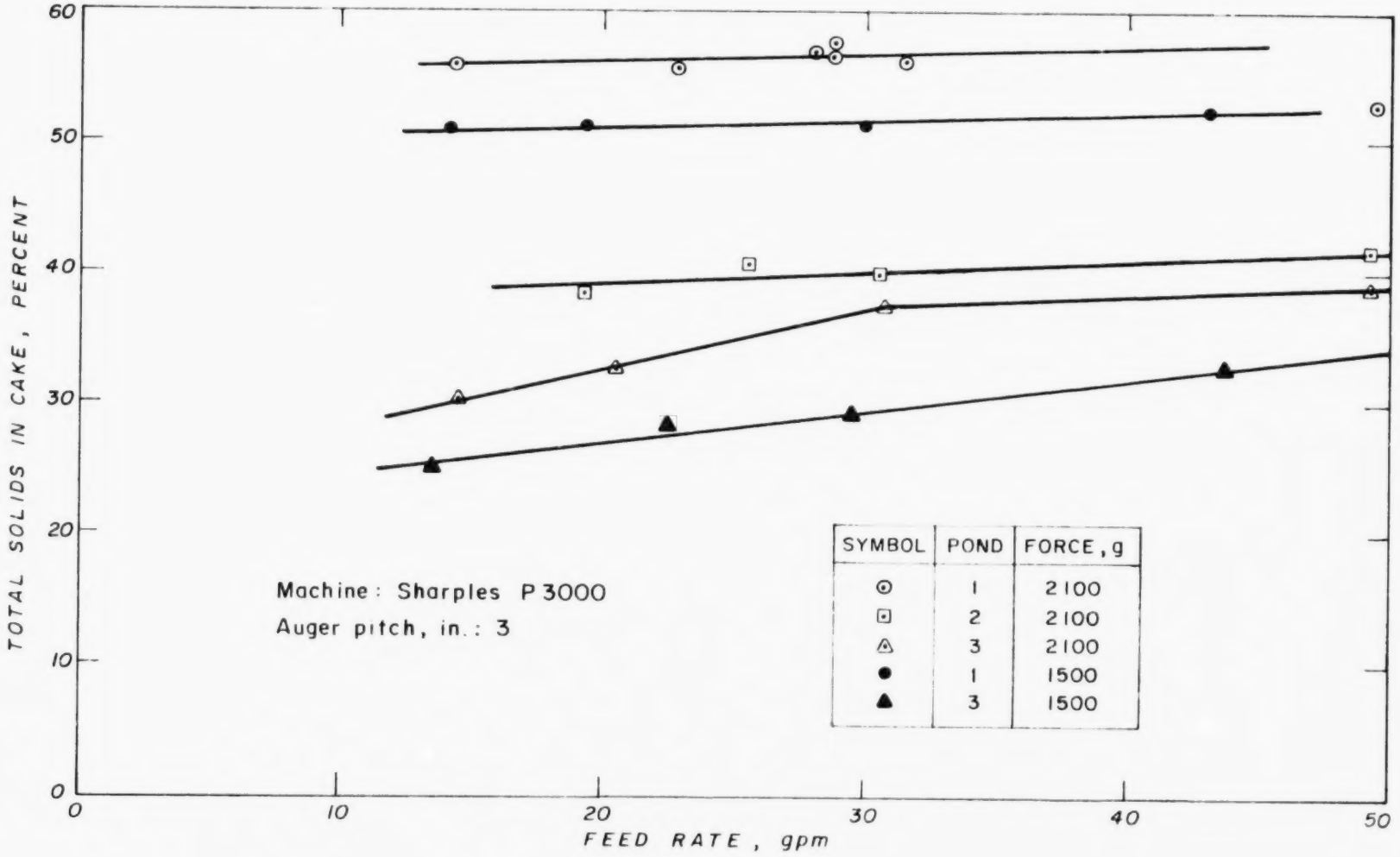



Figure 7-6

Dewatering at Flocculation pH of 11.0

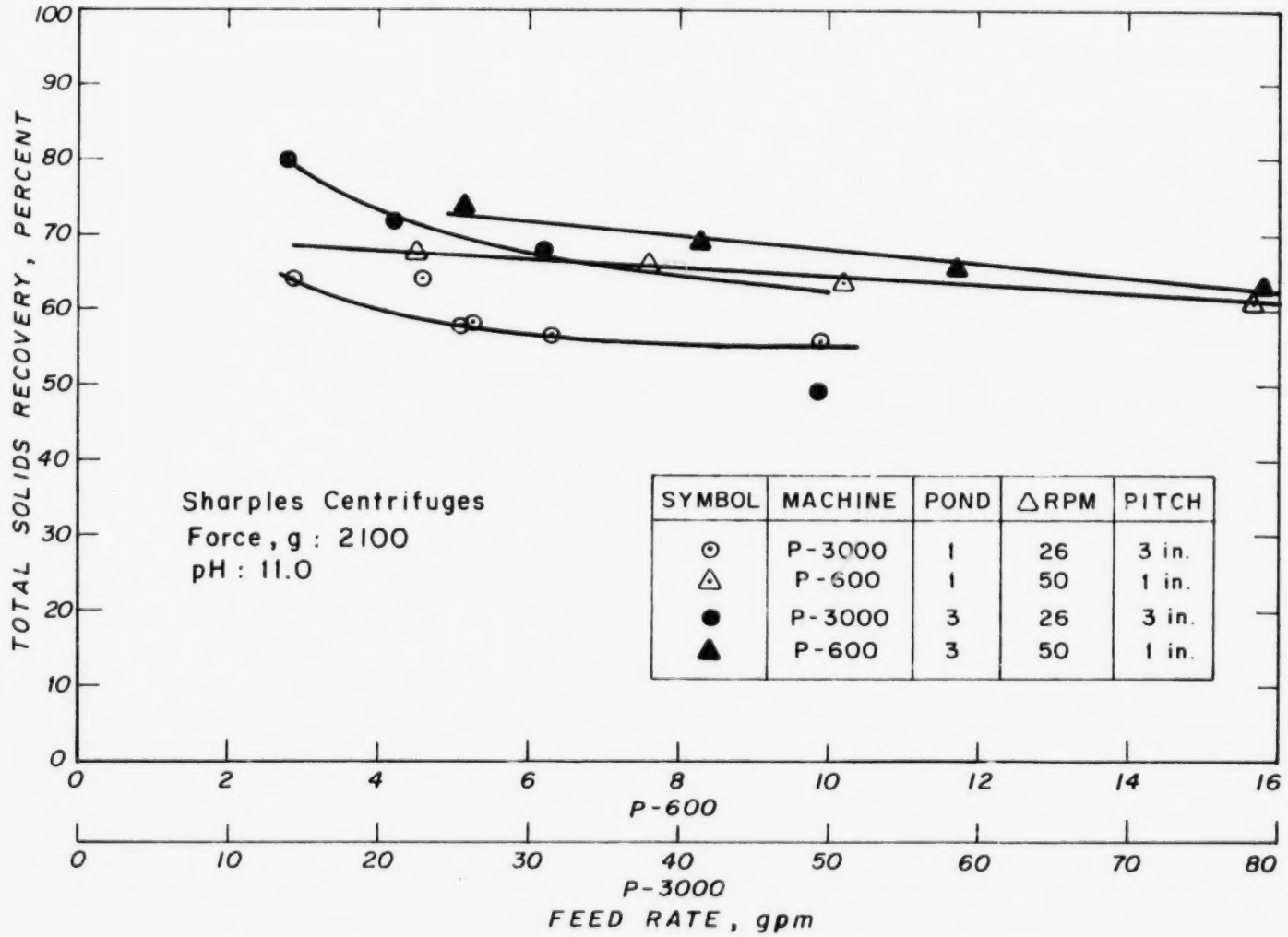



Figure 7-7 Comparison of Centrifuges for Solids Recovery

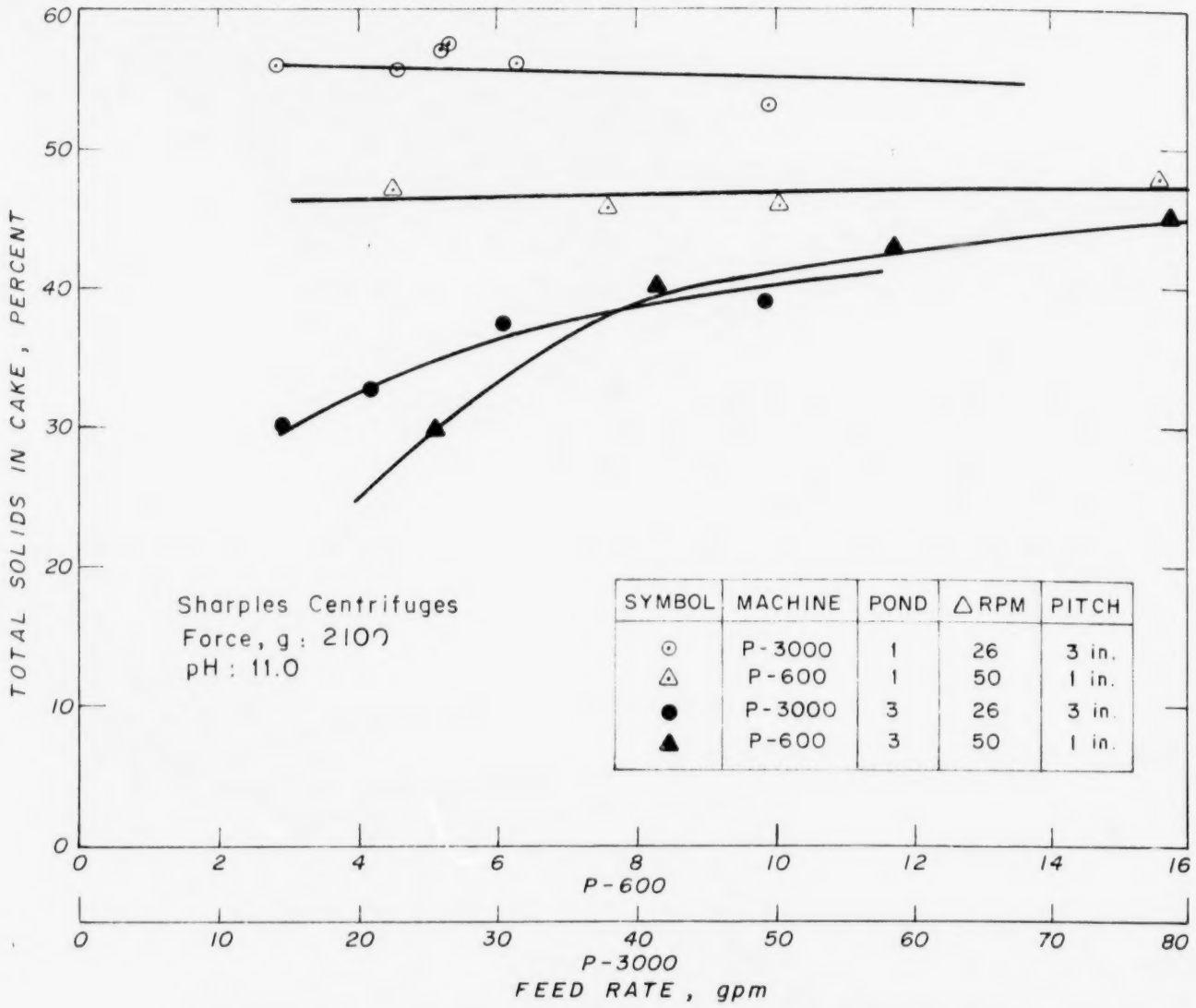

TOTAL SOLIDS IN CAKE, PERCENT
60
50
40
30
20
10
0
Sharples Centrifuges
Force, g : 2100
pH : 11.0
SYMBOL | MACHINE | POND | △RPM | PITCH
⊙ | P-3000 | 1 | 26 | 3 in.
△ | P-600 | 1 | 50 | 1 in.
● | P-3000 | 3 | 26 | 3 in.
▲ | P-600 | 3 | 50 | 1 in.
0 2 4 6 8 10 12 14 16
P-600
0 10 20 30 40 50 60 70 80
P-3000
FEED RATE, gpm

Figure 7-8 Comparison of Centrifuges for Dewatering

removes solids from the centrifuge bowl. The studies on second stage operation (Section VIII) give an indication of the importance of proper auger speed. A high auger speed relative to the bowl results in excess water being removed with the settled solids. If the auger speed is too low, there will be an insufficient rate of solids removal, but the solids that are removed will be relatively dry.

For the two machines tested, the smaller machine with a higher auger speed produced a higher total solids recovery but a lower cake dryness than the larger machine. If the auger speeds of the two machines were comparable, it is anticipated that solids recovery and cake dryness would also be comparable.

Lime Dose or pH

As discussed under "constituent separations", the effect of lime dose or pH could not be established with absolute certainty because other variables were changed simultaneously. As lime dose or pH in the primary tank increased, it appeared that solids recovery also increased.

## SUMMARY

The calcium carbonate in primary lime sludge can be preferentially separated from the other constituents in the sludge by wet classification in a centrifuge. For a flocculation pH of 11.0 and 11.5 approximately 90 percent of the calcium carbonate was recovered while rejecting 50 to 75 percent of the other solid constituents. At a flocculation pH of 10.2 the calcium carbonate recovery was about 75 percent when rejecting 60 percent of the other materials. The calcium carbonate-rich cake was simultaneously dewatered during wet classification to a total solids content of 42 to 57 percent.

To attain a high recovery of calcium carbonate during wet classification and simultaneously dewater the captured solids, the centrifuge must be designed and operated to separate the constituents. The best results were obtained with a low liquid depth in the bowl (pond setting of 1). Increasing centrifugal force improved the dewatering substantially and the calcium carbonate recovery slightly. A centrifugal force of 2100 g appears to be adequate for wet classification and dewatering.

Another variable in centrifuge operation is the auger which removes solids from the centrifuge bowl. While a study of auger pitch and speed was not conducted in this study, the study of second stage centrifuge operation indicates that the auger shape and speed are important variables. For an operating installation flexibility in adjusting auger speed is considered desirable to attain the optimum recovery and concentration of calcium carbonate under any given set of conditions.

The centrifuge can be operated over a wide range of feed rates with only a small effect on the separation. An increase in feed rate caused a minor decrease in calcium carbonate recovery; however, only a small change in separation occurred for a twofold change in feed rate. Feed rate had only a minor effect on solids recovery and no significant effect on dewatering over a 3:1 range of feed rates.

In summary, the tests showed that a solid bowl centrifuge can adequately wet classify and dewater lime sludge.

## SECTION VIII

## CENTRATE PROCESSING BY CENTRIFUGING

In the wet classification of primary sludge, calcium carbonate is concentrated for recalcining and reuse as lime. The other constituents in the sludge are preferentially rejected to the centrate. A second centrifuge stage can be used to recover and dewater the rejected solids from the wet classification. The centrate from the second centrifuge stage can then be recycled to the primary treatment section of the plant. To avoid a large buildup of solids in the primary treatment section, the recycle stream must have a low solids content. Therefore, a high recovery of solids in the second stage is necessary.

If the recovered solids are to be incinerated, the water content of the solids must be as low as possible. Efficient dewatering will minimize the fuel requirement for incineration. Consequently, in the second centrifuge stage both the recovery and the dewatering of the solids should be as nearly complete as possible.

The solids to be captured in the second centrifuge stage are the slow settling solids which were rejected in the first centrifuge stage. Different conditions must therefore be used in the second stage than in the first stage. Efficient capture requires a high centrifugal force and the addition of a polymer to improve the settling characteristics. The factors affecting the capture and dewatering of solids are discussed in this section.

The centrate from the wet classification of primary sludge was accumulated in a storage tank and fed to a centrifuge to study the capture and dewatering of solids in first stage centrate. Most of the study was made with Sharples centrifuges, either the P600 or the P3000. A few runs were made with a 6-inch Bird unit. Samples of feed, centrate and cake were analyzed and recoveries calculated as described in Section VII. Results of all the centrifuge runs on first stage centrate are presented in Table 8-1.

### CONSTITUENT CLASSIFICATION

In wet classification calcium carbonate is preferentially captured, and the other solid constituents are preferentially rejected to the centrate. In the second centrifuge stage a high capture of solids with little or no discrimination between constituents is desirable. The test results (Table 8-1) showed that some classification of constituents seems to be unavoidable, although the degree of separation was much less than for the first stage. The highest

# Table 8-1

## Constituent Separation During Centrifuging of First Stage Centrate

| Run | Machine | Process Parameters: Centrifugal force per Bowl, g's | Process Parameters: Angel speed, differential rpm | Process Parameters: Angel pitch, in. | Process Parameters: Feed number | Process Parameters: Feed rate, gpm | Process Parameters: Flocculation time, min | Process Parameters: Polymer type[a] | Process Parameters: Polymer dose, lb/ton[b] | Centrifuge Performance: Dewatering, Total solids in cake, percent | Centrifuge Performance: Recoveries in Cake, weight percent: Total solids | Calcium carbonate | Magnesium hydroxide | Hydroxyapatite | Ferric hydroxide | Organic matter | Other materials |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 2-21 | Sharples P-600 | 3050 | 17 | 1 | 4 | 2.4 | 11.5 | 72S | 6.5 | 16.1 | 69 | ~2 | ~4 | 69 | | 63 | 52 |
| 2-34 | | | 24 | | 3-1-2 | 3.4 | | 2A2 | 3.3 | 13.8 | 87 | 99 | 79 | ~7 | | 78 | 99 |
| 2-41 | | 2100 | 16 | | | 2.0 | | | 5.5 | 13.4 | 86 | 93 | 100 | ~0 | | 71 | ~100 |
| 2-52[c] | | | 24 | | | 3.4 | | | 3.3 | 13.7 | 81 | ~100 | ~0 | ~1 | | 71 | 81 |
| 3-1 | | 3050 | 16 | | 5 | 2.0 | 11.0 | | ~5 | 13.5 | ~6 | ~100 | ~5 | 79 | ~0 | 71 | 74 |
| 3-2 | | | 16 | | | 2.5 | | | 6.6 | 15.9 | 78 | ~100 | 7 | ~0 | 78 | 67 | 79 |
| 3-4 | Bird 5-in. bowl | 2050 | na[d] | na[d] | na[d] | 2.2 | 10.2 | | 6.2 | 12.1 | 89 | ~100 | 99 | 94 | | 79 | 89 |
| 3-6 | | | | | | 2.2 | | | 7.3 | 11.5 | 95 | ~100 | ~100 | 99 | | 83 | 86 |
| 3-12 | | | | | | 2.9 | | | 9.3 | 14.6 | 87 | ~100 | ~100 | ~5 | | 69 | 89 |
| 3-16 | | | | | | 3.0 | | | 7.8 | 15.7 | 81 | ~100 | ~100 | ~0 | | 62 | 86 |
| 3-11[c] | Sharples P-600 | 2100 | 20 | 1 | 4 | 3.3 | | | 5.6 | 14.6 | 86 | 91 | ~5 | 94 | 96 | 80 | (24) |
| Average of all runs | | | | | | | | | | 14.2 | 82 | 97 | ~9 | ~3 | | 73 | 77 |

[a] Allied polymer colloid 72S or ICI America polymer Atlasep 2A2.

[b] Pounds polymer per ton of dry solids in feed.

[c] Feed consisted of 11 percent sludge and 89 percent first stage centrate.

[d] Information not available.

recoveries were obtained for calcium carbonate, while the [illegible] were for organic matter and the constituents destroyed [illegible]. However, the extent of classification between constituents was relatively small. Consequently, variations in operating conditions had only a minor affect on the relative recovery of individual constituents, and the major concern was the dewatering and capture of total solids.

## SOLIDS RECOVERY AND DEWATERING

As mentioned previously a high efficiency in both capture and dewatering of solids is desirable. Efficient capture will maximize the recycle flow to the sewage treatment facilities, and efficient dewatering will minimize the heat requirements for incineration. Since a polymer is required to flocculate the small particles and improve their settling characteristics, one of the most important considerations is the effect of centrifuge operating conditions on the amount of polymer required to attain a given separation. While a research study to define quantitatively the effect of each variable is beyond the scope of this study, sufficient data were obtained to establish the approximate efficiency of solids capture and dewatering and the relative importance of the operating variables.

### Polymer Type

Several polymers are available for flocculating solids in a centrifuge. Since the type of polymer has a marked effect on the degree of separation, a study of various polymers was made to identify one that could be used satisfactorily.

Several polymers were given cursory examination in the laboratory with a beaker test. The test consisted of adding the polymer to first stage centrate in a beaker and gently stirring the mixture. The effectiveness of the polymer was judged by visual observation of the floc.

The results of the beaker tests are shown in Table 8-2 for a centrate from a primary flocculation pH of 10.2. A second series of tests was run for a primary flocculation pH of 11.5 but deleting Diamond Shamrock 930 polymer and adding American Cyanamid Magnafloc 835A and Calgon 2700. Similar results were obtained from both tests. In both cases ICI Atlasep 2A2 and Allied Colloid 725 produced the best flocs.

The two polymers that appeared best by the beaker test were further evaluated in centrifuge tests. The results are shown in Fig. 8-1. ICI Atlasep 2A2 gave the best results both in the beaker test and the centrifuge runs and was therefore used for all subsequent studies of the second centrifuge stage.

## Table 8-2

## Laboratory Beaker Test of Polymers for Second Stage Centrifuge Feed

| Polymer | Size of Floc[a] |
|---|---|
| ICI Atlasep 2A2 | large |
| Allied Colloid 725 | medium |
| Hercules 836.2 | small |
| ICI Atlasep 3A3 | very small |
| Allied Colloid 726 | very small |
| Dow A-23 | very small |
| Diamond Shamrock 930 | very small |

[a]Centrate used in the tests was produced from a primary flocculation pH of 10.2. Polymer dosage in each case was 5 lb per ton dry solids.

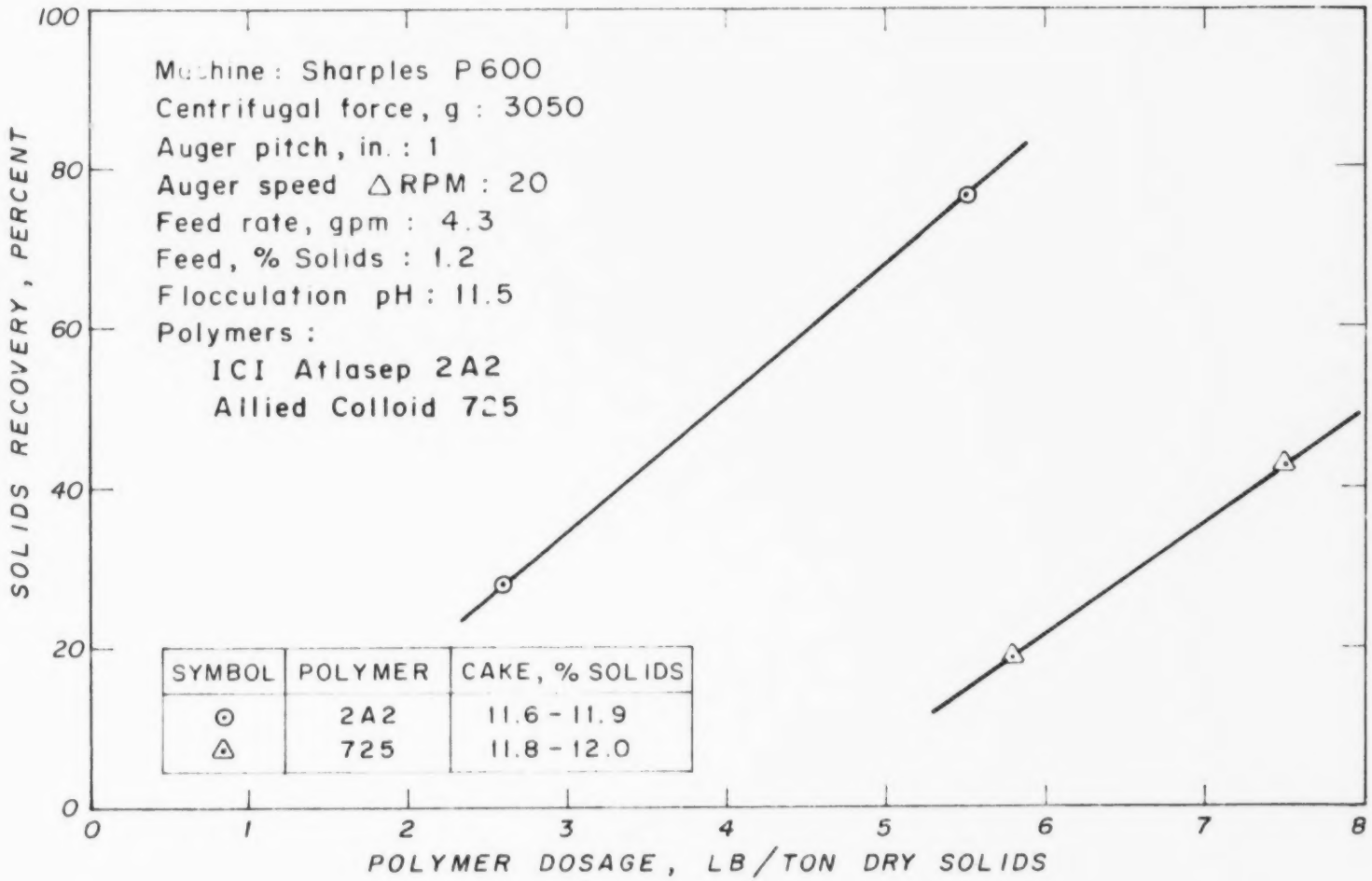



Figure 8-1

Effect of Polymer Type on Separation in Centrifuge

## Centrifuge Operating Variables

It was beyond the scope of the present study to establish the quantitative effect of the many interdependent variables in centrifuge operation. The study was restricted to obtaining data for selection of prototype equipment sizes and for establishing the areas where operating flexibility is necessary.

Auger Speed. Since the solids separated by centrifugal force are removed from the bowl with an auger, the auger speed relative to the bowl ($\Delta$ RPM) is one of the important operating variables. An example of the effect of auger speed is shown in Fig. 8-2. The polymer requirement to attain an 80 percent recovery of solids was reduced to about half by increasing the auger speed from 8 to 12 $\Delta$ PPM. However, the extent of dewatering was reduced by the increased auger speed. Apparently there is an optimum auger speed that gives the best balance between polymer usage and extent of dewatering for a given feed and set of operating conditions.

Pond Setting. Several factors are involved when the liquid depth is changed. Increasing the depth increases the liquid residence time, which should improve the seperation. Offsetting to an extent the increased residence time is a reduced centrifugal force at the liquid surface due to the fact that the surface is nearer the hub of the revolving bowl. Consequently, the average centrifugal force is reduced. Increasing pond depth also increases solids removal efficiency. As the depth is increased the solids are more easily "lifted over" the beach of the bowl by the auger. For the second centrifuge stage Fig. 8-3 shows that increasing the liquid depth reduced the required polymer dosage to attain a given recovery, but also reduced the solids content of the cake.

On occasions difficulty was experienced in getting the solids removed from the bowl at a pond setting of less than four, particularly at low polymer dosages. The solids would tend to build up in the bowl and then "dump", making an unstable operating condition.

Centrifugal Force. Recovery would normally be expected to improve with an increase in centrifugal force, with the only limitation being the equipment and operating cost relative to the improvement in degree of recovery. As shown in Fig. 8-4, the solids recovery is improved as the centrifugal force is increased. However, on the Sharples P3000 machine no improvement in recovery was obtained by increasing the centrifugal force above 2,100 g. The cake dryness did continue to improve, but the improvement from 2,100 to 3,200 g was small. Apparently there is little advantage in applying a centrifugal force greater than 2,100 g.

Feed Rate. In the centrifuge test runs shown in Fig. 8-5 the feed rate was varied from 14 to 29 gpm with no significant change in polymer requirement or percent of solids recovered. There was an apparent slight increase in cake

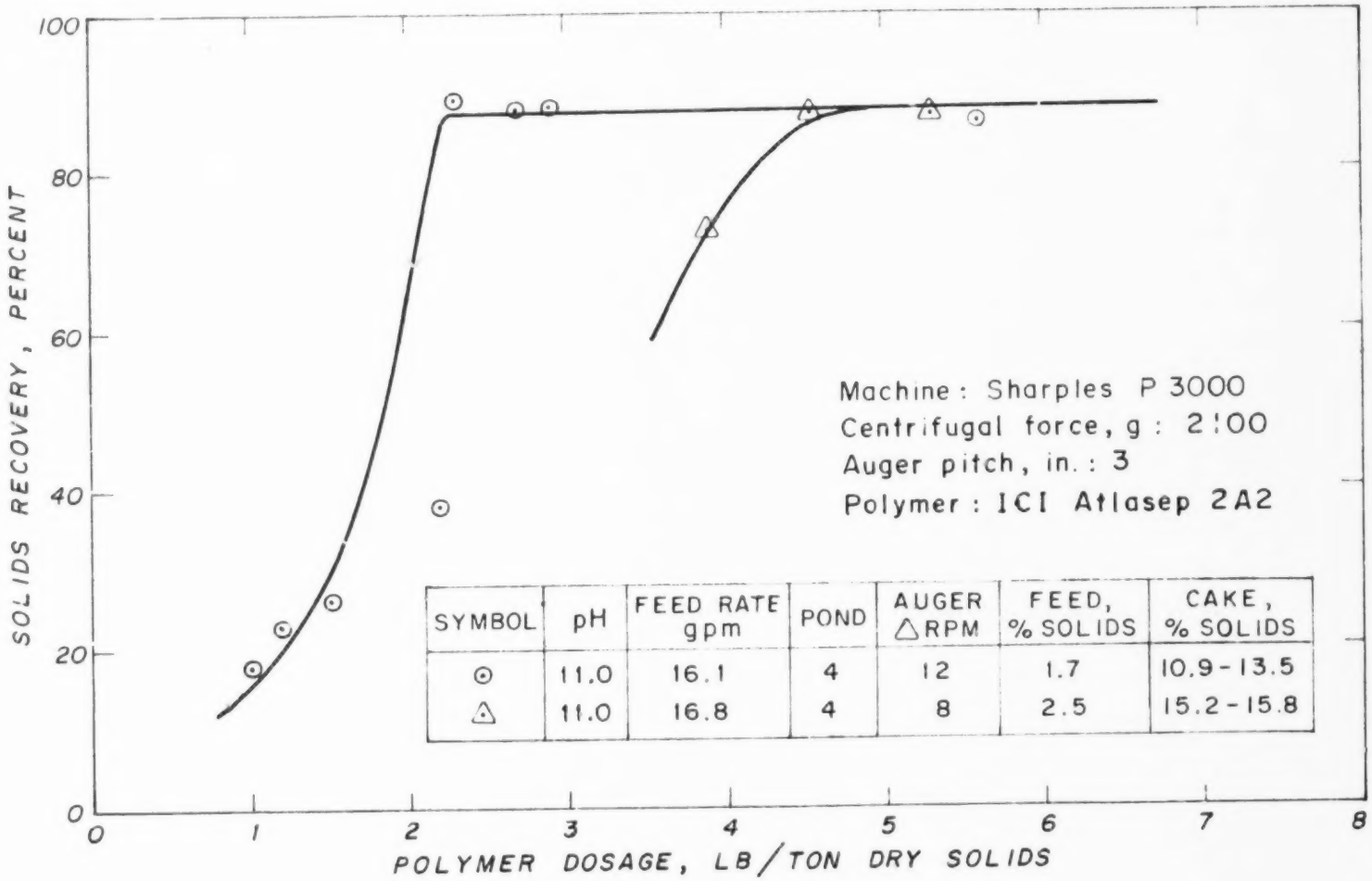



Figure 8-2

Effect of Auger Speed on Polymer Requirement

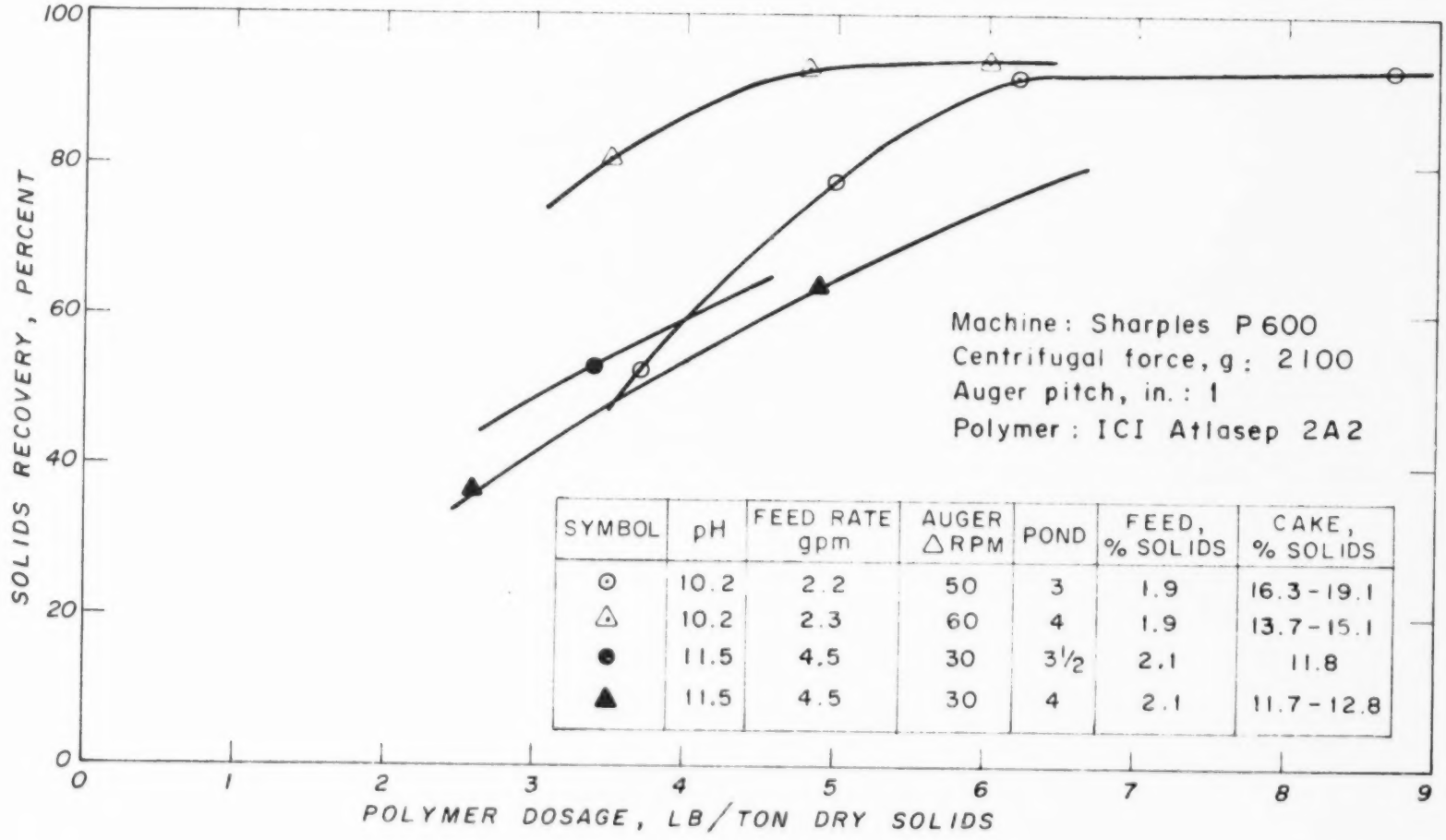

SOLIDS RECOVERY, PERCENT
POLYMER DOSAGE, LB/TON DRY SOLIDS
Machine: Sharples P 600
Centrifugal force, g: 2100
Auger pitch, in.: 1
Polymer: ICI Atlasep 2A2
SYMBOL | pH | FEED RATE gpm | AUGER △RPM | POND | FEED, % SOLIDS | CAKE, % SOLIDS
⊙ | 10.2 | 2.2 | 50 | 3 | 1.9 | 16.3-19.1
△ | 10.2 | 2.3 | 60 | 4 | 1.9 | 13.7-15.1
● | 11.5 | 4.5 | 30 | 3½ | 2.1 | 11.8
▲ | 11.5 | 4.5 | 30 | 4 | 2.1 | 11.7-12.8

Figure 8-3

Effect of Pond Setting on Recovery of Solids

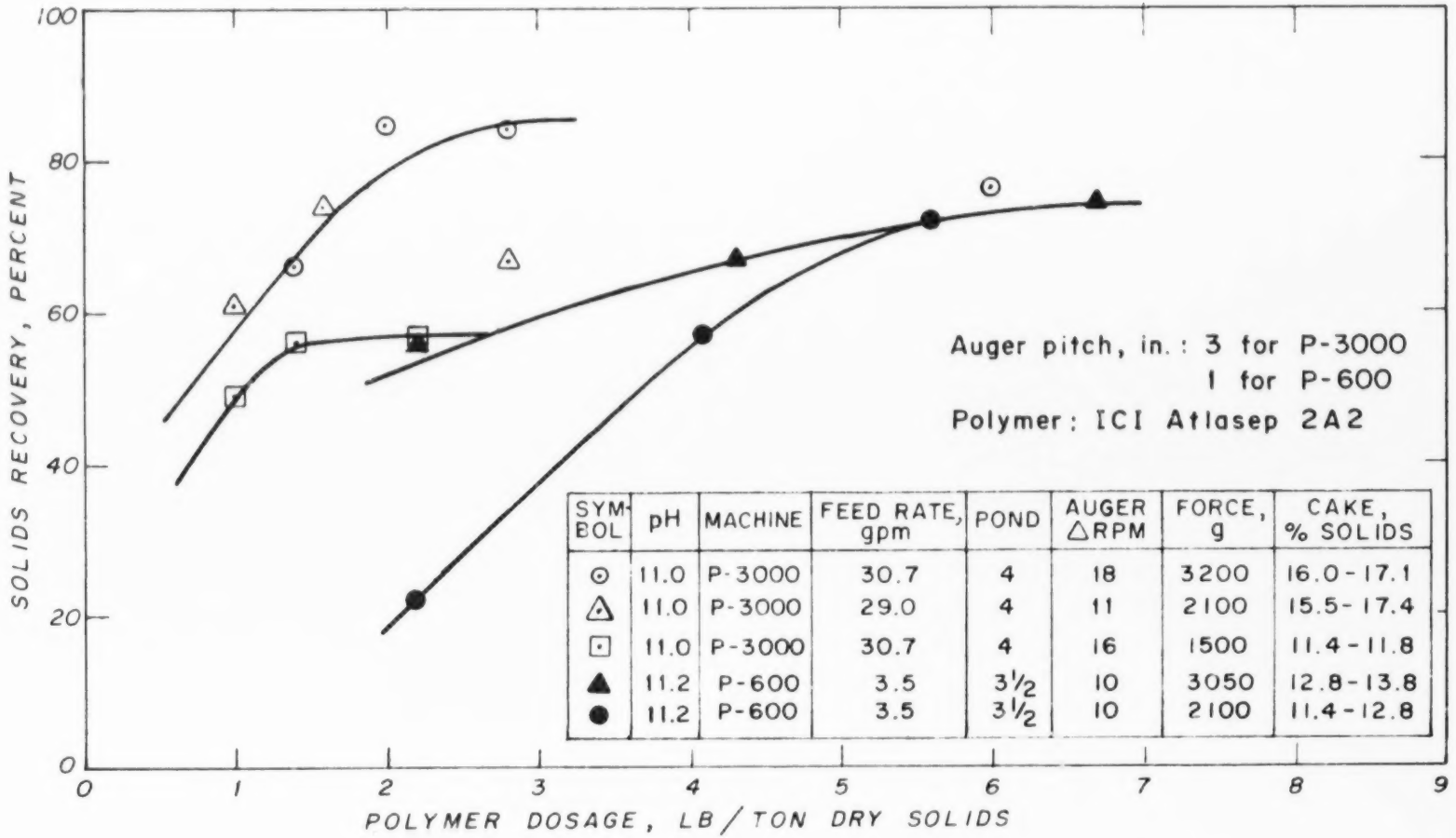

SOLIDS RECOVERY, PERCENT
POLYMER DOSAGE, LB/TON DRY SOLIDS
Auger pitch, in.: 3 for P-3000
1 for P-600
Polymer: ICI Atlasep 2A2
SYMBOL | pH | MACHINE | FEED RATE, gpm | POND | AUGER △RPM | FORCE, g | CAKE, % SOLIDS
⊙ | 11.0 | P-3000 | 30.7 | 4 | 18 | 3200 | 16.0-17.1
△ | 11.0 | P-3000 | 29.0 | 4 | 11 | 2100 | 15.5-17.4
⊡ | 11.0 | P-3000 | 30.7 | 4 | 16 | 1500 | 11.4-11.8
▲ | 11.2 | P-600 | 3.5 | 3½ | 10 | 3050 | 12.8-13.8
● | 11.2 | P-600 | 3.5 | 3½ | 10 | 2100 | 11.4-12.8

Figure 8-4 Effect of Centrifugal Force on Recovery

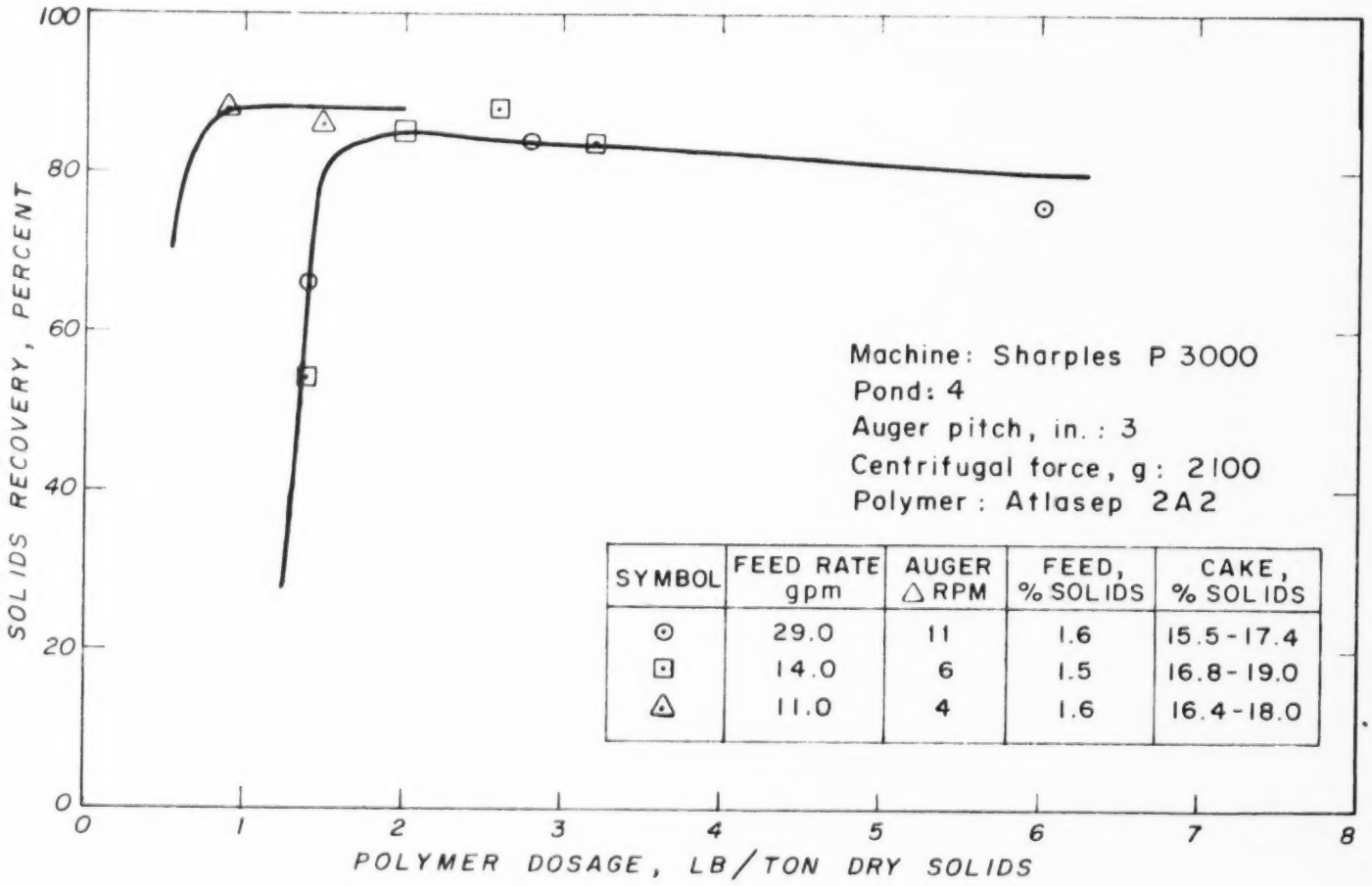

SOLIDS RECOVERY, PERCENT
POLYMER DOSAGE, LB/TON DRY SOLIDS
Machine: Sharples P 3000
Pond: 4
Auger pitch, in.: 3
Centrifugal force, g: 2100
Polymer: Atlasep 2A2
SYMBOL | FEED RATE gpm | AUGER △ RPM | FEED, % SOLIDS | CAKE, % SOLIDS
⊙ | 29.0 | 11 | 1.6 | 15.5-17.4
⊡ | 14.0 | 6 | 1.5 | 16.8-19.0
△ | 11.0 | 4 | 1.6 | 16.4-18.0

Figure 8-5 Effect of Feed Rate on Recovery

water content at the higher feed rate. At the low feed rate of 11 gpm, about a third of the maximum rate used, solids recovery was slighly better, and the polymer requirement was reduced. Fig. 8-5 shows that a wide range of feed rates can be used without changing the efficiency to any great extent.

For the runs at different feed rates the auger speed was also varied to maintain a relatively constant ratio of feed rate to auger speed. Since the concentration of solids in the feed was essentially constant, the amount of solids removed per revolution of the auger was also maintained constant, thereby essentially eliminating the effect of auger speed on cake dryness.

Comparison of Machines. Two different sizes of Sharples centrifuges were used for the experimental work. To estimate the effect of machine size on solids recovery, all of the data for the two machines for similar feed rates and at similar centrifugal forces were plotted in Fig. 8-6. The figure shows that the larger machine, the P3000, gave better solids recovery with a lower polymer dose than the smaller machine, the P600.

Effect of Feed pH on Centrifuge Performance

Since pH affects the composition of the solids in sludge, it was postulated that the ease of separation of the solids from the liquid might also be affected. A study was therefore made of the effect of varying the pH of the feed to the centrifuge for both the first and second centrifuge stages. The study consisted of two phases. The first phase evaluated the effect of changes in the pH of flocculation, which over the course of the experimental work was varied from 10.2 to 11.5. The second phase used sludge from a constant flocculation pH and evaluated the effect of adjusting the pH of the feed to the first and second centrifuge stages.

Flocculation pH. It is difficult to identify the effect of flocculation pH, because the study extended over a period of several months and encompassed several changes in operating conditions. There is also the problem that the centrifuge work was performed to obtain design data rather than to define the effect of all operating variables such as pH. Consequently, changes were made in centrifuge sizes and operating conditions without regard to flocculation pH.

All of the data on centrifuge operation were reviewed, and specific runs were selected which appeared to offer a valid basis for comparison. Since no two runs at different flocculation pH levels were made under the same operating conditions, a direct comparison is impossible. However, it was established that solids recovery varies with the solids removal rate, which is a function of auger speed and solids feed rate. A comparison was therefore made between runs at different flocculation pH levels but a constant ratio of solids feed rate to auger speed. The results are shown in Fig. 8-7 for different levels of centrifugal force. The figure does not show any relationship between floccu-

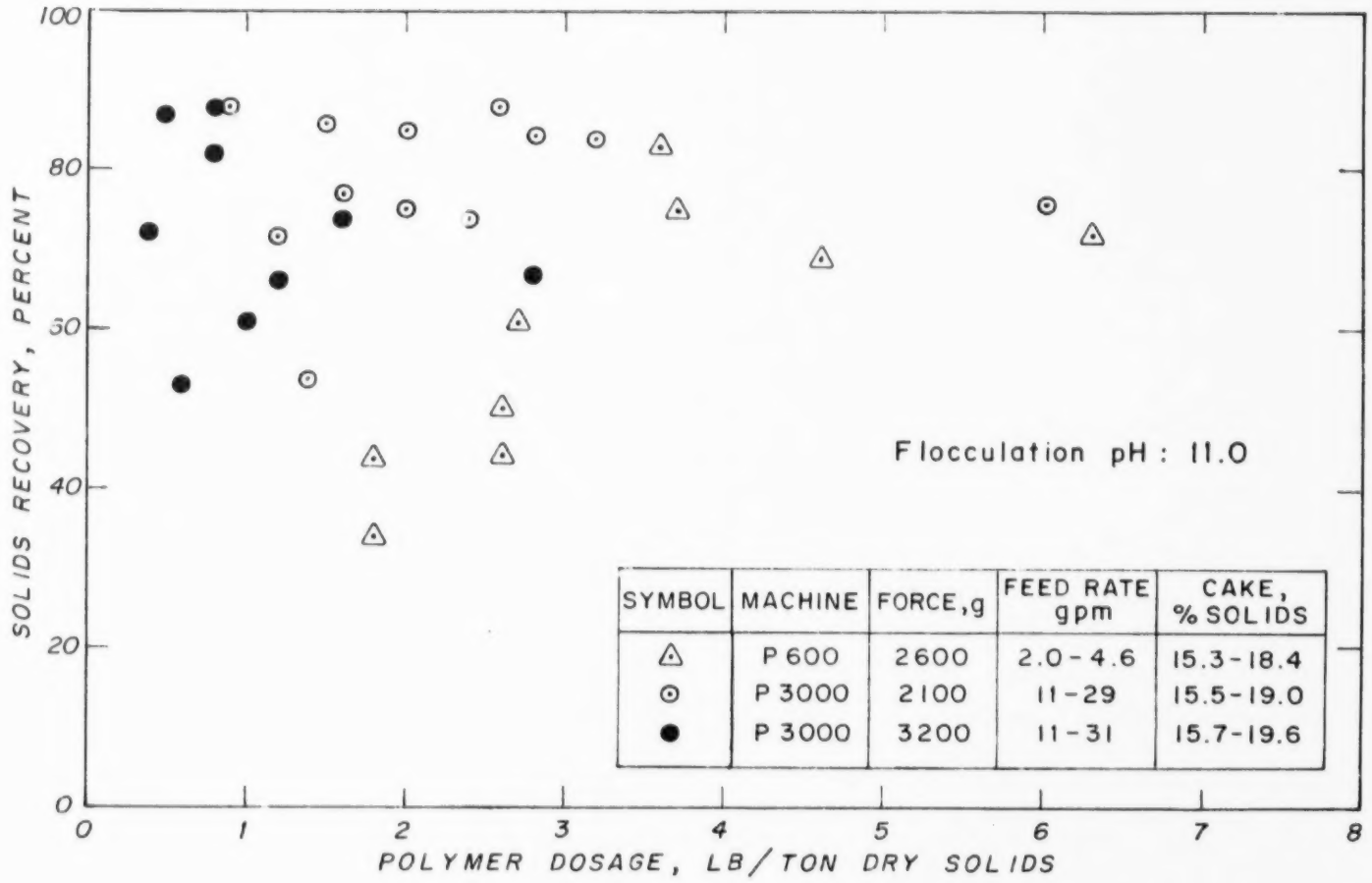
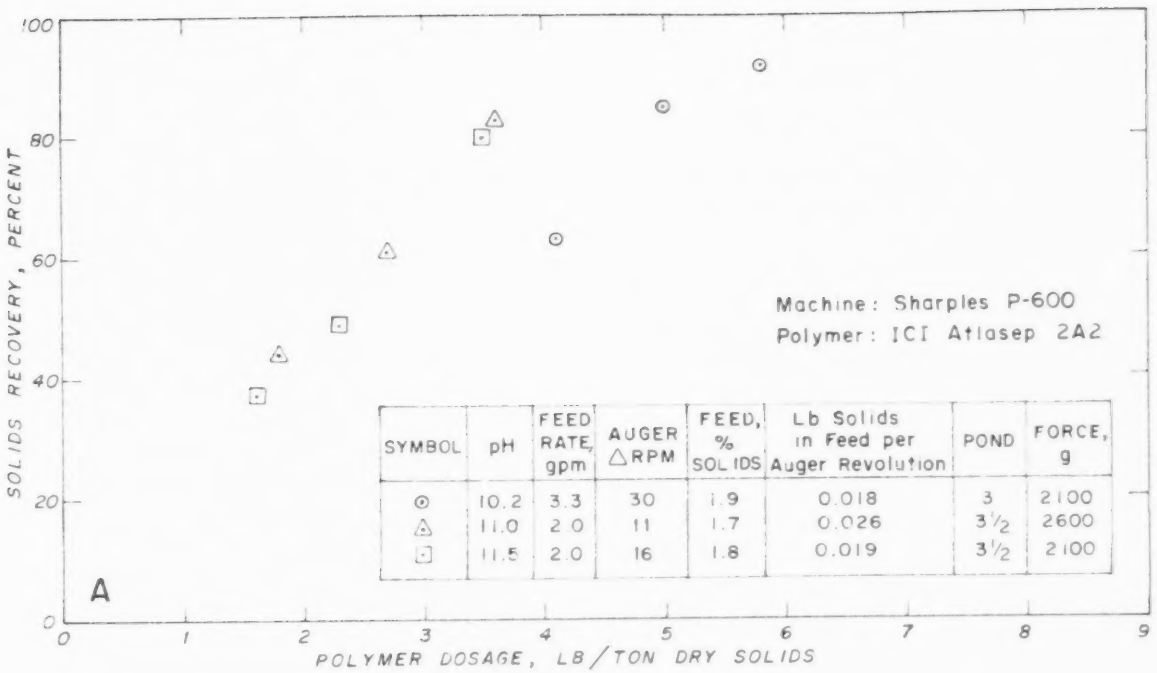
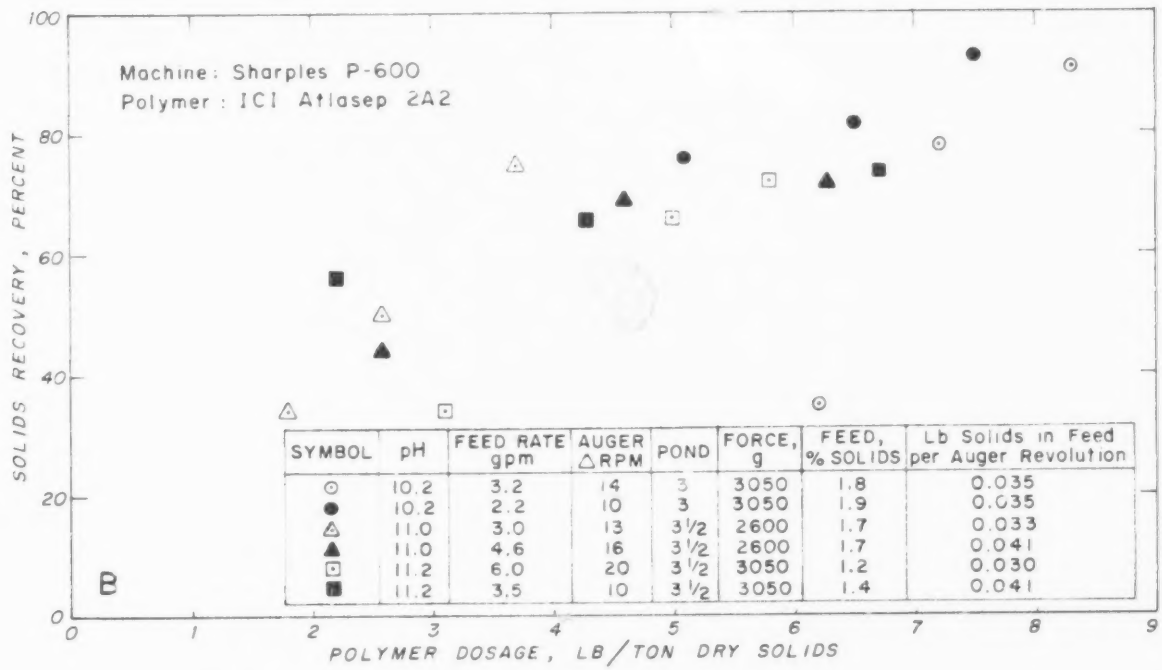



Figure 8-6 Comparison of Two Sharples Machines for Second Stage Operation




| SYMBOL | pH | FEED RATE, gpm | AUGER △RPM | FEED, % SOLIDS | Lb Solids in Feed per Auger Revolution | POND | FORCE, g |
|---|---|---|---|---|---|---|---|
| ⊙ | 10.2 | 3.3 | 30 | 1.9 | 0.018 | 3 | 2100 |
| △ | 11.0 | 2.0 | 11 | 1.7 | 0.026 | 3½ | 2600 |
| ⊡ | 11.5 | 2.0 | 16 | 1.8 | 0.019 | 3½ | 2100 |




| SYMBOL | pH | FEED RATE gpm | AUGER △RPM | POND | FORCE, g | FEED, % SOLIDS | Lb Solids in Feed per Auger Revolution |
|---|---|---|---|---|---|---|---|
| ⊙ | 10.2 | 3.2 | 14 | 3 | 3050 | 1.8 | 0.035 |
| ● | 10.2 | 2.2 | 10 | 3 | 3050 | 1.9 | 0.035 |
| △ | 11.0 | 3.0 | 13 | 3½ | 2600 | 1.7 | 0.033 |
| ▲ | 11.0 | 4.6 | 16 | 3½ | 2600 | 1.7 | 0.041 |
| ⊡ | 11.2 | 6.0 | 20 | 3½ | 3050 | 1.2 | 0.030 |
| ■ | 11.2 | 3.5 | 10 | 3½ | 3050 | 1.4 | 0.041 |

Figure 8-7A, B

Effect of Flocculation pH on Solids Recovery

lation pH and solids recovery. The observed differences in recovery could well have been due to other differences in operating conditions.

Flocculation pH values greater than 11.0 had a detrimental effect on second stage cake water content, as shown in Table 8-3.

Adjustment of Centrifuge Feed pH. Some lime sludge processing flow sheets include pH adjustment of the sludge with carbon dioxide ("recarbonation") prior to sludge processing. The purpose of recarbonation is to dissolve the magnesium hydroxide fraction which may hinder dewatering if left in the sludge [21].

Sludge generated at a flocculation pH of 11.0 was adjusted to various levels to assess the effects on centrifuge performance. On some occasions a pH drop was accomplished by bacterial action in the primary, on most occasions carbon dioxide was used. When the pH was raised, lime was used.

Solids recovery in the first centrifuge stage varied between 55 and 58 percent. Cake solids varied from 53 to 60 percent, with the wettest cake from a feed pH of 10.5. The observed variations are probably all within the range of experimental error, and it was concluded that the effect of pH on first stage separation is negligible.

The data for the second stage centrifuge are presented in Fig. 8-8. Several combinations of first stage feed pH and second stage feed pH are indicated.

The following observations can be drawn from the data:

1. Adjusting the feed to the first stage centrifuge to a pH greater than 11.0 makes dewatering in the second stage more difficult. A much higher polymer dosage was required for solids capture for first stage feeds adjusted to pH 11.5 than for lower pH sludges.

2. Adjustment of the feed to the second stage centrifuge to a pH greater than 11.0 seems to have no detrimental or beneficial effect.

3. In general, there appears to be little or no benefit derived from recarbonation of the first or second stage feed sludge.

Effect of pH Adjustment on Composition of Solids. Adjusting the pH of the centrifuge feeds was expected to change the solubility of the constituents and, consequently, change the composition of the solids. To estimate the extent of change in solids composition the soluble calcium, magnesium, and phosphorus in the feed and centrate from several first stage centrifuge runs were determined. The results are given in Figs. 8-9, 8-10 and 8-11. The change in

## Table 8-3 Effect of Flocculation pH on Second Stage Cake Dryness

| pH | Total solids in cake, percent | |
|---|---|---|
| | Range | Median |
| 11.2 to 11.5 | 11.0 to 14.4 | 11.8 |
| 11.0 | 15.5 to 19.6 | 17.2 |
| 10.2 | 13.7 to 20.3 | 18.4 |

## Table 8-4 Effect of Adjusting pH of Centrifuge Feed on Solids Composition

| pH | Supernatant Liquid Composition, mg/l[a] | | | Percent of Solids Dissolved[b] | | | | | |
|---|---|---|---|---|---|---|---|---|---|
| | | | | Sludge[c] | | | Centrate[d] | | |
| | Calcium as $CaCO_3$ | Magnesium as $CaCO_3$ | Phosphorous as $PO_4$ | $CaCO_3$ | $Mg(OH)_2$ | $Ca_5OH(PO_4)_3$ | $CaCO_3$ | $Mg(OH)_2$ | $Ca_5OH(PO_4)_3$ |
| 11.5 | 440 | 10 | 0.04 | 0.7 | -0.4 | nil | 3.8 | -0.9 | nil |
| 11.0 | 290 | 30 | 0.07 | - | - | - | - | - | - |
| 10.5 | 220 | 50 | 0.10 | -0.4 | 0.1 | nil | -1.5 | 0.3 | nil |
| 10.0 | 180 | 80 | 0.17 | -0.6 | 1.1 | nil | -2.5 | 1.9 | nil |
| 9.5 | 180 | 120 | 0.38 | -0.6 | 1.9 | 0.01 | -2.5 | 4.2 | 0.01 |
| 9.0 | 230 | 200 | 1.16 | -0.3 | 3.7 | 0.04 | -1.2 | 8.1 | 0.05 |

[a] From trend lines shown in Figures 8-9, 8-10, and 8-11.

[b] Compared to material at pH 11.0. A negative value indicates material precipitated rather than dissolved.

[c] For a sludge containing 5 percent total solids, and solids consisting of 40% $CaCO_3$, 5% $Mg(OH)_2$ and 5% $Ca_5OH(PO_4)_3$.

[d] For a first stage centrate containing 2 percent total solids, and solids consisting of 20% $CaCO_3$, 6% $Mg(OH)_2$ and 16% $Ca_5OH(PO_4)_3$.

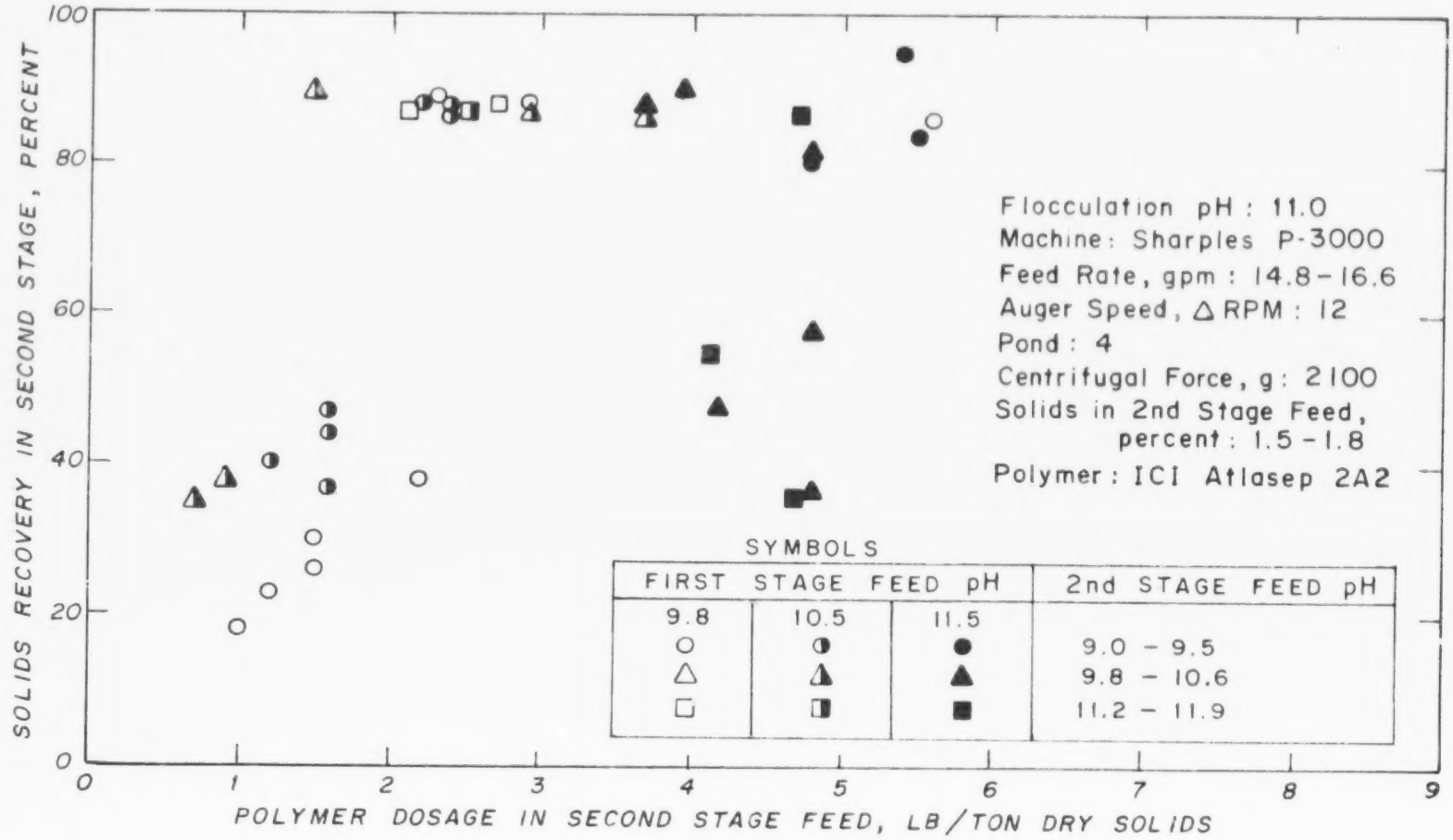

SOLIDS RECOVERY IN SECOND STAGE, PERCENT
POLYMER DOSAGE IN SECOND STAGE FEED, LB/TON DRY SOLIDS
Flocculation pH : 11.0
Machine: Sharples P-3000
Feed Rate, gpm : 14.8 - 16.6
Auger Speed, △ RPM : 12
Pond : 4
Centrifugal Force, g : 2100
Solids in 2nd Stage Feed, percent : 1.5 - 1.8
Polymer : ICI Atlasep 2A2
SYMBOLS
FIRST STAGE FEED pH
9.8
10.5
11.5
2nd STAGE FEED pH
9.0 - 9.5
9.8 - 10.6
11.2 - 11.9

Figure 8-8

Effect of Adjusting pH of Centrifuge Feed

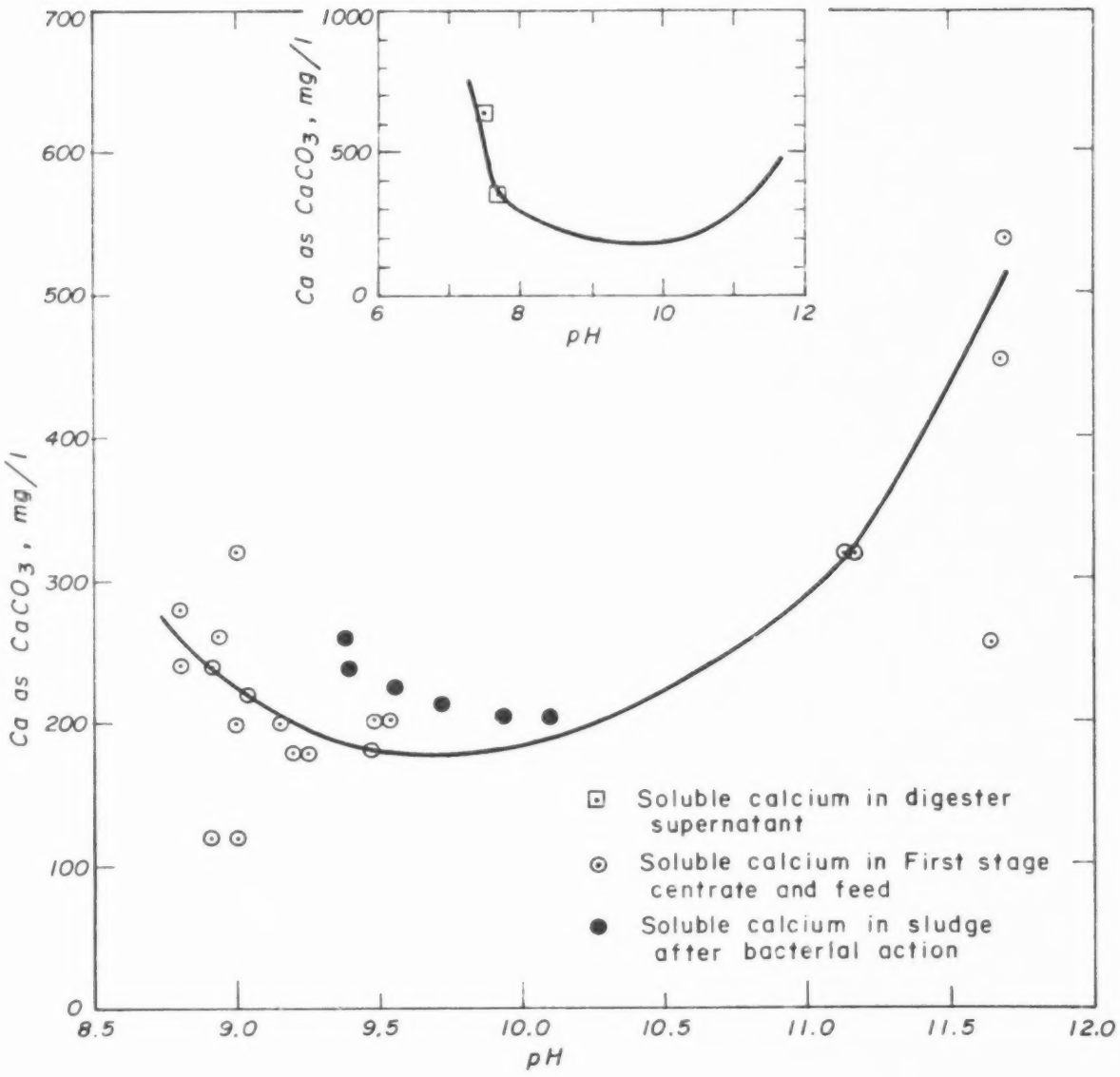

Ca as CaCO3, mg/l
pH
Soluble calcium in digester supernatant
Soluble calcium in First stage centrate and feed
Soluble calcium in sludge after bacterial action

Figure 8-9

Effect of pH on Solubility of Calcium

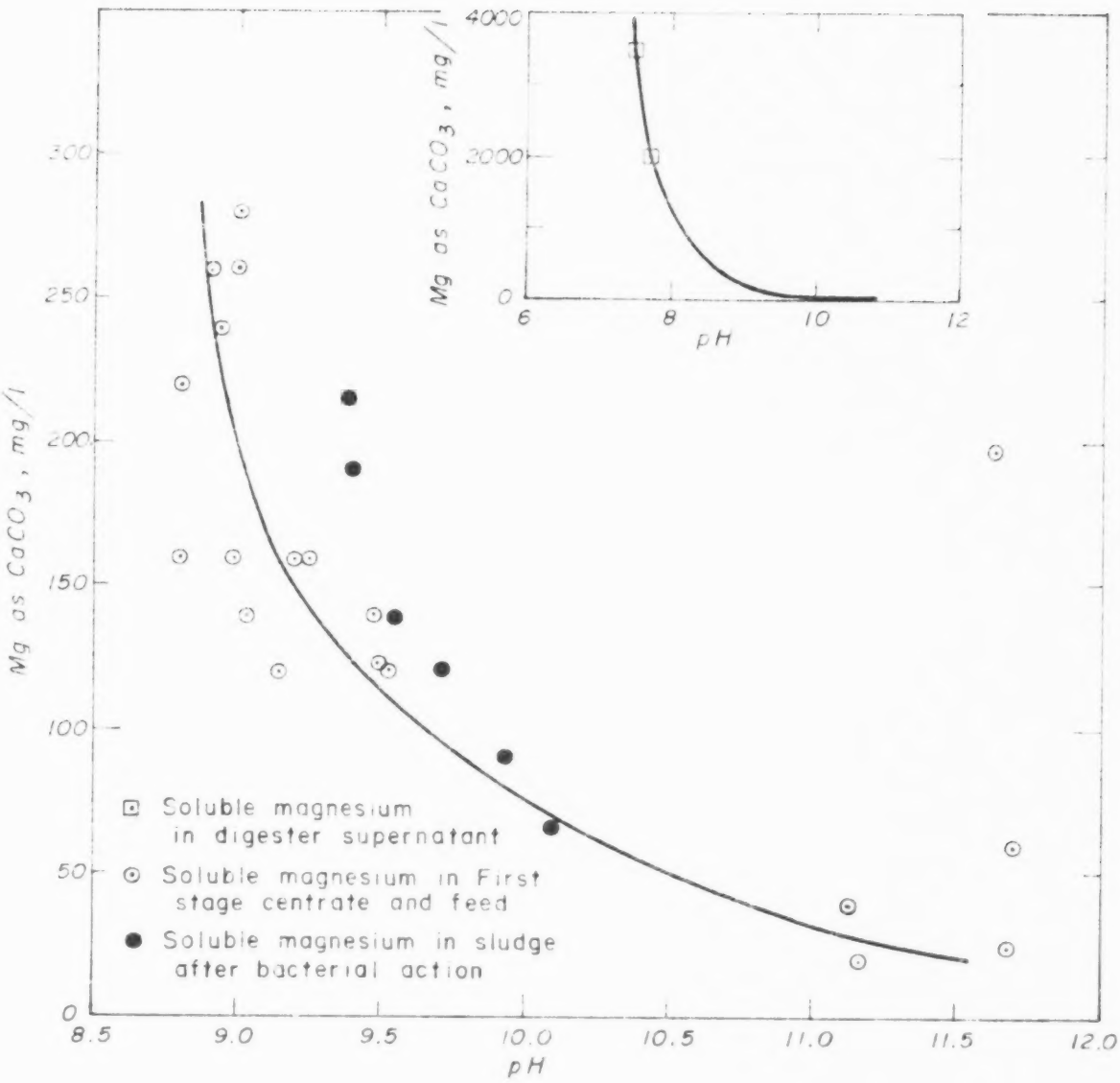



Figure 8-10

Effect of pH on Solubility of Magnesium

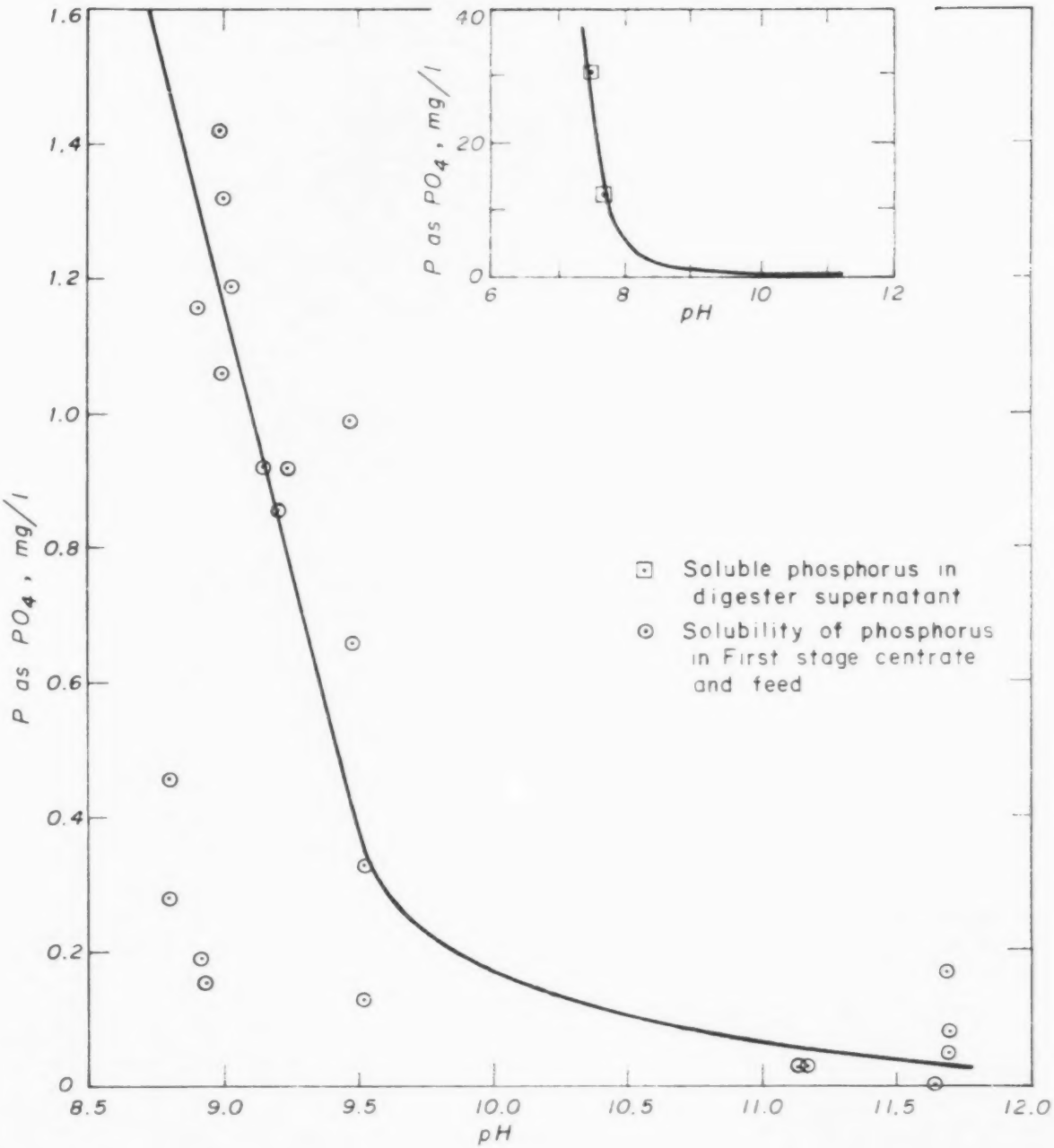



Figure 8-11

Effect of pH on Solubility of Phosphorus

solids composition was estimated from the curves. As shown in Table 8-4, the change in solids composition over the pH range of 9.0 to 11.5 was small. The largest change in centrate solids occurred in magnesium. An estimated 6 percent decrease in the magnesium content of the centrate solids occurred when the pH was reduced from 11.0 to 9.0.

The observed change in solubility of magnesium with pH is much less than would be expected for equilibrium conditions. The magnesium solubility can be estimated from solubility equilibrium[19]. The comparison between calculated and observed solubilities is shown in Table 8-5.

The large discrepancy between the calculated equilibrium conditions and the observed results is evidently due to the fact that equilibrium was not attained when adjusting pH. Evidently, a longer reaction time is required than was employed when making the pH adjustment. Reaction times varied, but usually were several hours.

In another experiment a sample of sludge was gently stirred in a series of beakers for different time periods. During that time the pH decreased as a result of bacterial action. The results are shown in Table 8-6 and plotted in Figs. 8-9 and 8-10.

Even though 13 hours elapsed while the pH was reduced from 10.1 to 9.4 the amount of dissolved magnesium was far below the equilibrium value. This again indicates a slow approach to equilibrium.

Also shown in Figs. 8-9, 8-10 and 8-11 are the levels for soluble calcium, magnesium and phosphorus in the supernatant from the experimental digesters described in Section IX. In this case, many days of reaction time was available and for the digester with liquid retention times of 15 days or greater nearly complete dissolution of magnesium was indicated. This indicates that the reaction time for solubilization is measured in days rather than hours.

In conclusion, perhaps the reason that sludge recarbonation (decreasing the pH) had little effect on sludge solids composition is that major changes in sludge constituents did not occur within the reaction times employed in the study.

Effect of Adding Sludge to Second Stage Feed

In an effort to reduce the polymer requirement, chemical primary sludge was added to the second stage feed. It was hoped that the fibrous solids in the sludge would assist the capture of second stage solids. The results are shown in Fig. 8-12. At a given polymer dosage, a higher solids recovery and a slightly denser cake were obtained when sludge was added to the feed. However, since no polymer is required to capture approximately 60 percent of the solids

## Table 8-5
## Comparison of Observed and Theoretical Magnesium Solubilities

| pH | Solubility of magnesium as calcium carbonate, mg/l | |
|---|---|---|
| | Calculated | Observed |
| 9.5 | 3,610 | 120 |
| 9.0 | 36,100 | 200 |

## Table 8-6
## Hardness Release Due to Bacterial Action

| pH | Time, hr | Solubility, mg/l | |
|---|---|---|---|
| | | Calcium as $CaCO_3$ | Magnesium as $CaCO_3$ |
| 10.1 | 0 | 206 | 66 |
| 9.9 | 2 | 206 | 92 |
| 9.7 | 4.3 | 214 | 122 |
| 9.6 | 6 | 226 | 140 |
| 9.4 | 11 | 240 | 192 |
| 9.4 | 13 | 260 | 216 |

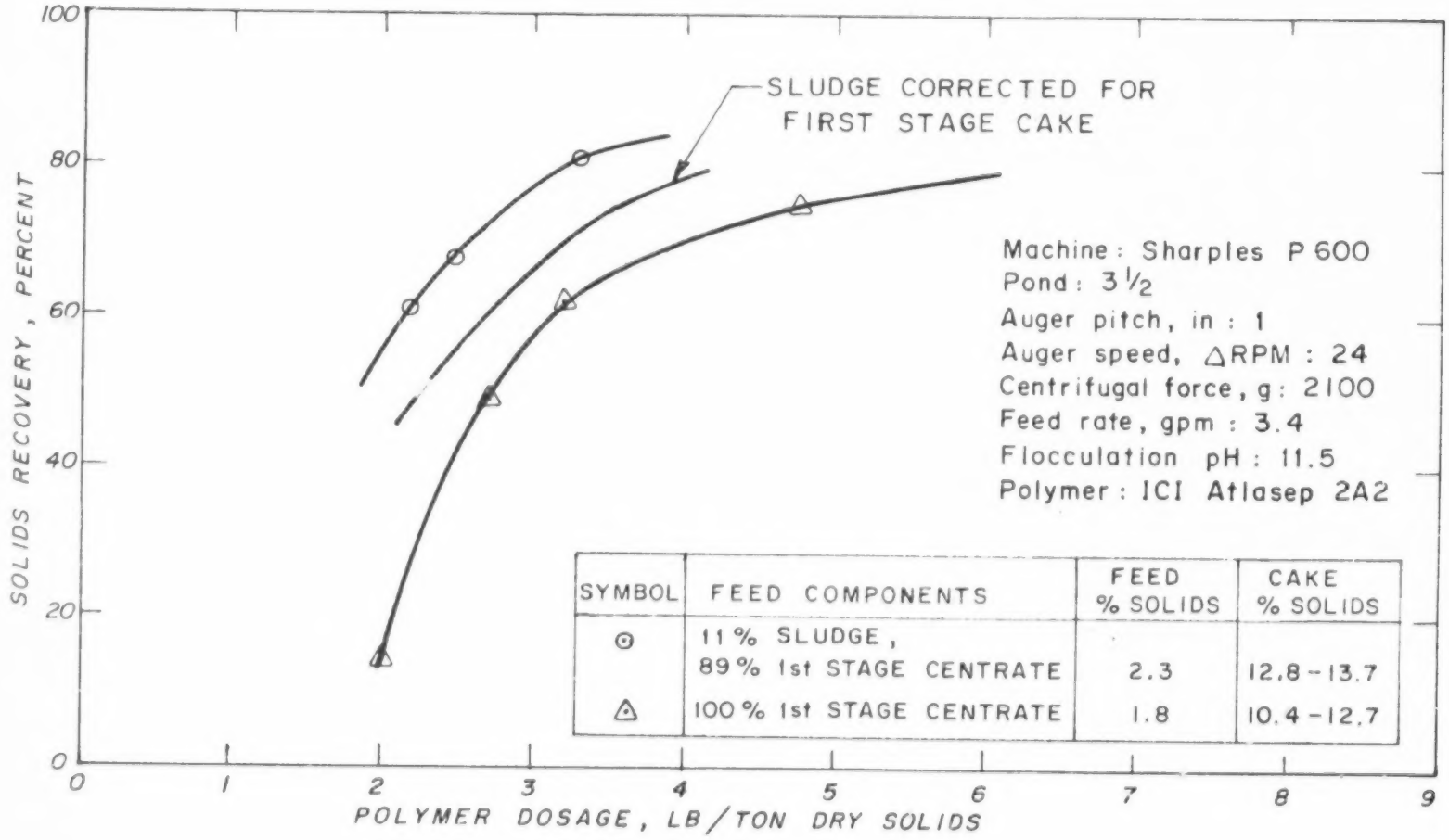



Figure 8-12

Effect of Adding Sludge to Second Stage Feed

in the raw sludge which was added to the centrate, the results should be corrected for the first stage solids. Correcting the recoveries and polymer dosages for those solids in the combined sludge which required no polymer gives the corrected curve in Fig 8-12. After correction the data still show a slight improvement in solids recovery by adding sludge to the second stage feed. However, the benefit is marginal.

## SUMMARY

The study of centrifuging first stage centrate demonstrates that a high recovery and satisfactory dewatering of the solids can be obtained. For a flocculation pH of 11.0 a recovery of 80 percent or more of the solids can be realized using a polymer dose of two pounds per ton of dry solids. Cake solids content greater than 15 percent can be achieved. With the proper operating conditions a polymer dose of two pounds per ton of solids should be adequate for capturing 80 percent of the solids for any flocculation pH between 10.2 and 11.5. However, dewatering of the solids was less efficient at a flocculation pH of 11.5 than at lower pH values.

An evaluation was made of the effect of recarbonation or lime addition to adjust the pH of either the chemical primary sludge or first stage centrate prior to centrifuging. The pH adjustment had little or no beneficial effect on solids recovery in the pH range tested. Raising the pH of the first stage feed above 11.0 increased the polymer requirement substantially, but at or below 11.0 adjustments in pH had at most a minor effect on the polymer requirement.

The polymer requirement for a given solids recovery was sensitive to certain machine configurations and operating conditions. Considerably less polymer was required with the Sharples P3000 centrifuge than with the smaller Sharples P600. Whether the difference was due to differences in size or in certain configurations, such as the auger pitch and shape, is not known.

One of the most important centrifuge operating variables is the rate of solids removal from the centrifuge bowl. For a given machine, the rate of solids removal is mainly a function of auger speed. The auger speed must be carefully adjusted to allow the rate of solids removal to match the rate of solids capture in the centrifuge. Consequently, a high degree of flexibility in adjusting auger speed is necessary.

The capacity of a centrifuge appears to be far more dependent on the rate of solids removal than the liquid flow rate. Consequently, satisfactory solids recovery can be obtained in a given unit over a relatively wide range of feed rates provided that the rate of solids removal is adjusted for differences in rate of solids intake.

As would be expected, increasing the centrifugal force improved the separation and decreased polymer requirement. However, increasing centrifugal force above 2,100 g produced marginal benefits.

Increasing liquid depth in the centrifuge bowl improved solids recovery but decreased the efficiency of dewatering. A pond setting of four was found to be optimum for the beach height of the machines used.

In summary, satisfactory solids recovery and dewatering of first stage centrate can be attained with a centrifuge. For an operating installation, flexibility in the rate of polymer addition and rate of solids removal from the centrifuge bowl will improve operating efficiency.

# SECTION IX

# CENTRATE PROCESSING BY ANAEROBIC STABILIZATION

Although incineration of the waste solids is planned for the initial stages of the CCCSD reclamation plant, incineration is not the only available alternative for later stages. Anaerobic digestion may be attractive, provided that suitable disposal sites are available. Moreover, since sludge digestion is a net energy producing process as opposed to an energy consuming process such as incineration, availability of fossil fuels may in the future establish digestion as a preferable solution to the sludge disposal problem.

Studies were conducted on a bench scale to evaluate anaerobic digestion as a means of processing the primary lime sludge solids in the first stage centrate from the solid bowl centrifuge. The centrate was thickened to two different solids levels. The low solids lime sludge was fed to the digesters at a TS concentration of 2.5 percent. The high solids tests were conducted at TS concentrations of 3.3 and 7.1 percent. Digester performance was evaluated at each level. Digesters were operated at different solids retention times (SRT) for the lime sludge feed, while a control digester was operated using raw primary (non-chemical) sludge feed as a comparison. Also, limited data were obtained from operation of a main plant digester which was fed both lime sludge and conventional primary sludge.

## OBJECTIVES OF STUDY

Because there is very little information available on the anaerobic stabilization of chemical sludges, the pilot study was conducted to evaluate operational parameters. Among these are:

1. Operational characteristics of digesters, such as pH, alkalinity, volatile acids and gas composition.

2. Digester performance, including gas production, volatile matter destruction, loading effects, and odor and drying characteristics of digested sludge.

3. Concentration ranges of soluble calcium, magnesium, and orthophosphate in the digesters; total calcium, magnesium and phosphorus in the digested sludge; and heavy metals, ammonia nitrogen and soluble sulfide levels in the digested sludge.

4. Corrosion and scaling evidence in circulation lines used in the experimental system.

## DESCRIPTION OF STUDY

The bench scale digester work was conducted using five-gallon glass bottles as the reactors. The reactors were kept at 38 C. The digesters were filled to a three-gallon level and kept well-mixed by recirculating the sludge with external pumps which were activated for approximately five minutes every half-hour by an electric timer control switch. Gas produced from the digesters was collected by the hydraulic displacement principle in separate calibrated bottles filled with a slightly acidic solution. Because the digesters were well mixed, solids retention time equaled hydraulic retention time. Initially three digesters were fed lime sludge at different retention times, and later a fourth was added. This gave operating data at four different retention times from 10 to 25 days. A control digester was fed raw primary sludge at a 15-day SRT.

### Analyses

The feed and digested sludges were periodically analyzed for pH, alkalinity, total and volatile solids, and soluble calcium, magnesium, and orthophosphate. Also a limited number of analyses were conducted for ammonia nitrogen, soluble sulfides, heavy metals, hexane extractable material, and total calcium, magnesium and phosphorus. The digester gas was analyzed for carbon dioxide by an Orsat apparatus, and this method was checked by chromatograph to verify the carbon dioxide and methane content of the gas.

### Performance Parameters

The experimental data were used to calculate the following parameters:

1. Percent volatile matter reduction.

2. Pounds of volatile matter fed per cubic foot of digester capacity per day.

3. Cubic feet of total gas and methane gas per pound of volatile matter fed (or destroyed).

### Chronological Description of Experiment

The first period of digester operations evaluated the performance when digesting first stage centrate of relatively low total solids (2.5 to 3.0 percent) containing 45 to 50 percent volatile solids. The digesters were operated for 47 days. Three digesters were fed lime sludge and operated at solids retention times of 15, 20, and 25 days. A control digester was run for three weeks at an SRT of 15 days, with a feed of about 2.5 percent total solids containing 68 percent volatile solids.

The second phase of experimental work lasted 40 days for the lime sludge digesters and 60 days for the control digester. During this second period the two lime sludge digesters that had been operating with an SRT of 15 and 25 days were fed sludge of 6 to 7 percent total solids of which 45 to 50 percent were volatile. A new lime sludge digester with a 10-day retention time was placed in operation on this same feed. The digester that had been operating at a 25-day SRT was fed lime sludge with 3 percent total solids in order to evaluate the effect of a slight increase in feed solids - possibly enough to overcome any near-washout conditions which may have been present during the previous runs with low feed solids. During the second phase the control digester was fed conventional sludge of 4 to 6 percent total solids of which 60 to 65 percent were volatile.

Additional analyses during this second period included ammonia nitrogen ($NH_3$-N); total calcium (Ca), magnesium (Mg) and phosphorus (P); heavy metals; and hexane extractable materials. At the completion of the study, all circulating lines and pumps were examined for corrosion and deposits.

## PERFORMANCE HISTORY AND OPERATIONAL SUMMARY

The operational results presented in this section will compare the stabilization of lime sludge and conventional sludge. Of particular importance is the evaluation of the ability of the anaerobic digestion process to stabilize lime sludge.

### Low Feed Solids Operation

To initiate the digestion process, seed sludge from the treatment plant digesters was placed in the reaction bottles, and a progressive feeding schedule was commenced. The initial operation of the bench-scale lime and control digesters was characterized by slow adaptation of the methane fermenting bacteria, resulting in high volatile acids for the first two weeks. After acclimation of the methane bacteria, volatile acids decreased from around 2000 ppm to under 100 ppm (principally as acetic acid). This reduction was accompanied by a simultaneous increase in gas production. The initial high volatile acids levels were probably the result of shocks both in type of feed and in temperature imposed upon the seed sludge.

Values of pH reached equilibrium at about 7.7 for the lime sludge digesters and 7.0 for the control digester. The soluble alkalinity (as $CaCO_3$) was generally between 3500 and 4000 ppm for the lime sludge digesters, while the control digester was usually between 2700 and 3000 ppm. Time plots of pH, alkalinity and volatile acids are shown in the appendix.

It is interesting to note that all of the lime sludge digesters showed a sharp drop in gas production simultaneously after about two weeks of

operation. Gas production curves are shown in the appendix. The drop could have been caused by a toxic contaminant in the common lime sludge feed supply. A contributing cause could have been that the concentration of organisms was so low due to the dilute feed that resistance to toxic compounds was poor[13]. The low solids content in the lime sludge digesters was possibly near the borderline of washing out at the lower solids retention times.

The control digester performed well for about three weeks on the lower solids feed. Thicker feed was then introduced, and the sudden shock caused a drop in performance until the digester could equilibrate again.

Table 9-1 shows a summary of the important performance parameters for the low feed solids period. The following observations can be drawn from comparisons between the control and lime sludge digesters:

1. Better volatile matter reduction was observed in the control digester than in the lime sludge digesters.

2. Higher methane content occurred in the lime sludge digester gas than in the control digester gas. This was due primarily to the pH difference between the two types of digesters; $CO_2$ was absorbed more readily in the lime sludge digesters.

3. Methane production per lb of volatile matter destroyed was about the same for control and lime sludge digesters.

4. Methane production per lb volatile matter fed was greater in the control digester than in the lime sludge digesters.

It was also noted that there was no detectable hydrogen sulfide ($H_2S$) odor in the lime sludge digester gas, while the control digester gas did have a slight $H_2S$ odor. The higher pH in the lime sludge digesters caused more sulfide to be in the ionized form in solution, whereas the lower pH in the control digester permitted more favorable equilibrium to hydrogen sulfide gas.

It is possible to observe from comparisons among the lime sludge digesters the following:

1. Greater volatile matter destruction and more methane produced per lb of volatile matter destroyed (or fed) in the lower rate (20 and 25 day) lime digesters than in the unit having an SRT of 15 days.

2. Essentially no difference in the methane produced per lb of volatile matter fed in the 20 and 25 day digesters. Overall performance of these two units was quite similar.

# Table 9-1

## Digestion Performance Parameters

| Feed sludge type and solids retention time | Loading lb VM per cu ft per day | Typical pH range | Digester TS,[b] wt percent | Overall VM destruction percent | Stable VM destruction,[c] percent | Cu ft gas per lb VM destroyed[d] | Cu ft $CH_4$ per lb VM destroyed[d] | Cu ft $CH_4$ per lb VM fed[d] | Percent $CO_2$ in gas[d] |
|---|---|---|---|---|---|---|---|---|---|
| | Low feed solids operation | | | | | | | | |
| Lime, 15 day | .05[a] | 7.6-7.8 | 2.2 | 34.2 | 34.9 | 12.3 | 11.3 | 3.86 | 8.4 |
| Lime, 20 day | .04[a] | 7.6-7.8 | 2.0 | 40.4 | 45.9 | 13.2 | 12.1 | 4.84 | 8.4 |
| Lime, 25 day | .03[a] | 7.6-7.8 | 2.0 | 44.8 | 44.8 | 13.8 | 12.6 | 4.66 | 8.9 |
| Raw (control), 15 day | .07[a] | 6.8-7.2 | 1.5 | 49.6 | 49.6 | 16.9 | 12.3 | 6.02 | 27.1 |
| | High feed solids operation | | | | | | | | |
| Lime, 10 day | .20[e] | 7.4-7.5 | 5.9 | 33.6 | 35.6 | 13.1 | 11.7 | 3.93 | 15.0 |
| Lime, 15 day | .14[e] | 7.4-7.5 | 5.6 | 38.3 | 41.8 | 12.9 | 10.9 | 4.17 | 15.5 |
| Lime, 20 day | .10[e] | 7.4-7.6 | 5.5 | 38.8 | 38.8 | 13.3 | 11.3 | 4.38 | 15.0 |
| Lime, 25 day | .08[e] | 7.4-7.6 | 2.7 | 41.8 | 41.8 | 15.1 | 13.3 | 5.55 | 12.3 |
| Raw (control), 15 day | .14[e] | 7.1-7.2 | 2.5 | 54.4 | 58.1 | 13.9 | 9.8 | 5.29 | 30.0 |

[a] Based on lime feed of 2.7 percent total and 45 percent volatile solids; and control feed of 2.5 percent total and 68 percent volatile solids.

[b] Values near end of runs.

[c] Stable periods shown on gas production curves in Appendix.

[d] Average values.

[e] Based on lime feed of 7.1 percent total and 47 percent volatile solids, for 10, 15, and 20 day SRT lime digesters; and lime feed of 3.3 percent total and 45 percent volatile solids for 25 day SRT digester; and control feed of 5.3 percent total and 63 percent volatile solids.

The soluble calcium, magnesium and orthophosphate analyses are listed for both the digested and feed sludges in Table 9-2. From these data one can see that:

1. Soluble orthophosphate is lower in the control digester effluent than in the feed. This is postulated to be because the feed sludge pH is lower than the digester pH, making possible the precipitation of phosphate under anaerobic conditions.

2. Lime sludge digestion releases a slight amount of soluble orthophosphate. Since anaerobic conditions reduce the pH of lime sludge the solubility of phosphate is increased. The amount released, however, is far less than the amount that could potentially be released if it were possible to dissolve all of the precipitated phosphorus. For instance, the sludge composition data in Section VI would indicate a value of total phosphorus (as orthophosphate) of about 750 mg/l for feed solids at 2.5 percent TS. Thus, less than 1.5 percent of the available orthophosphate was dissolved in the lime sludge digester.

3. Magnesium is released to a high concentration in the lime digesters. The concentration of soluble magnesium in the lime digesters was 2000 ppm as calcium carbonate ($CaCO_3$). Literature ranges for the effects of magnesium cation concentration on digester organisms are: stimulatory, 75-150 ppm as Mg (300-600 as $CaCO_3$); moderately inhibitory, 1000-1500 ppm as Mg (4150-6250 as $CaCO_3$); and strongly inhibitory, over 3000 ppm as Mg (12,500 ppm as $CaCO_3$)[12]. The concentration of magnesium in the lime sludge digester would therefore be between the stimulatory and moderately inhibitory ranges. The presence of other elements which could act either synergistically or antagonistically is unknown. Again, the sludge data in Section VI would indicate a total value of magnesium (as $CaCO_3$) of 1700 mg/l for a 2.5 percent TS feed. Thus, essentially all of the magnesium precipitate is dissolved in the digester.

4. No significant increase in soluble calcium or magnesium occurs in the control digester operation.

High Feed Solids Operation

The lime and control feed sludges were further thickened in an attempt to build a higher concentration of organisms in the digesters and possibly achieve more stable operation, especially in the lime sludge digesters.

## Table 9-2

## Feed and Digested Sludge Characteristics

| Sludge type and SRT | Total solids,[a] wt percent | Volatile solids,[a] wt percent | pH range | Alkalinity,[b] mg/l, as $CaCO_3$ | Volatile acids,[b] mg/l | Soluble Ca,[c] mg/l as $CaCO_3$ | Soluble Mg,[c] mg/l as $CaCO_3$ | Soluble P,[c] mg/l as $O\text{-}PO_4$ | Total Ca,[d] mg/l as $CaCO_3$ | Total Mg,[d] mg/l as $CaCO_3$ | Total P,[d] mg/l as $PO_4$ | Grease,[d] mg/l | Ammonia N, mg/l | Soluble sulfide,[d] mg/l |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Lime feed | 2.5–5.0 | 45–55 | 7.8–9.4 | 650–1000 | 511–650 | 110–560 | 380–1960 | 3.4–9.9 | 29,200 | 8503 | 624 | 9344 | 126 | .1 |
| Control feed | 2.5–6.1 | 60–70 | 5.3–6.5 | 600–1200 | 600–1300 | 300–400 | 420–450 | 117–123 | 4,000 | 1498 | 459 | 7616 | 190 | .1 |
| | Digested sludges, low feed solids period | | | | | | | | | | | | | |
| Lime, 15 day | 2.2 | 41 | 7.6–7.8 | 3400–4200 | 100–750 | 350 | 2000 | 11.6 | – | – | – | – | – | – |
| Lime, 20 day | 2.0 | 42 | 7.6–7.8 | 3500–4500 | 150–400 | 330 | 2000 | 11.4 | – | – | – | – | – | – |
| Lime, 25 day | 2.0 | 42 | 7.6–7.8 | 3500–4500 | 60–500 | 370 | 2200 | 14.6 | – | – | – | – | – | – |
| Control, 15 day | 1.5 | 62 | 6.8–7.2 | 2750–3100 | 100–1000 | 340 | 300 | 50.8 | – | – | – | – | – | – |
| | Digested sludges, high feed solids period | | | | | | | | | | | | | |
| Lime, 10 day | 5.9 | 36 | 7.4–7.5 | 4200–5300 | 850–1200 | 650 | 3450 | 31.4 | 19,500 | 6314 | 1291 | 3300 | 550 | .1 |
| Lime, 15 day | 5.6 | 35 | 7.4–7.5 | 4000–5500 | 350–700 | 600 | 3500 | 29.7 | 30,450 | 3860 | 4791 | 3674 | 600 | .1 |
| Lime, 20 day | 5.5 | 35 | 7.4–7.6 | 4500–5400 | 350–550 | 580 | 3500 | 29.7 | 29,400 | 3261 | 4798 | 2184 | 520 | .1 |
| Lime, 25 day | 2.7 | 37 | 7.4–7.6 | 3700–4350 | 52–250 | 620 | 2200 | 23.1 | 13,650 | 1946 | 1829 | 1352 | 300 | .1 |
| Control, 15 day | 2.5 | 47 | 7.1–7.2 | 3000–3600 | 50–150 | 390 | 330 | 48.2 | 8,052 | 1510 | 474 | 2211 | 580 | .1 |

[a] Values near ends of runs.

[b] Typical range during stable operation.

[c] Average value near end of run.

[d] Spot analyses only near end of high feed solids run, insufficient analyses run to be considered typical.

## Table 9-3

## Digester Sludge Heavy Metals Analyses at End of High Feed Solids Run

| Digester type and SRT | mg/Kg of dry total solids | | | | | | mg/Kg of dry fixed solids | | | | | | mg/l | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | Cr | Cu | Pb | Zn | Cd | Ni | Cr | Cu | Pb | Zn | Cd | Ni | Cr | Cu | Pb | Zn | Cd | Ni |
| Lime, 15 day | 70 | 356 | 239 | 522 | 12 | 67 | 109 | 556 | 373 | 815 | 19 | 104 | 4 | 21 | 14 | 30 | 1 | 4 |
| Control, 15 day | 78 | 398 | 343 | 681 | 15 | 663 | 159 | 812 | 700 | 1395 | 31 | 1353 | 3 | 13 | 11 | 23 | 1 | 22 |

The pH decreased from around 7.7 for the previous low lime feed solids period to around 7.5 for the high solids period in all the lime sludge digesters. The control digester pH showed no significant change from the previous low solids period. Alkalinities increased in the lime sludge digesters to around 5500 ppm as $CaCO_3$ in the 10, 15 and 20-day digesters, and to around 4400 ppm in the 25-day digester (still on "low" solids feed). The control digester alkalinity rose slightly to around 3400 ppm. Operating characteristics for all digesters during the high feed solids runs are shown in the appendix.

Performance parameters are shown in Table 9-1 for the high feed solids period. The following observations can be drawn from comparisons between the lime sludge and control digesters:

1. Better reduction of volatile matter was obtained in the control digester than in the lime sludge digester.

2. Again, higher methane content was found in the lime sludge digester gas than in the control digester gas.

3. Lime sludge digesters produced more methane per lb of volatile matter destroyed than did the control digester.

4. The control digester produced more methane per lb of volatile matter fed than did the 10, 15, and 20-day SRT lime sludge digesters, but produced slightly less than did the 25-day lime sludge digester.

From comparisons among the lime sludge digesters it is possible to conclude the following:

1. Percentage volatile matter destruction became progressively better as solids retention time increased.

2. The lime sludge digester with a SRT of 25 days produced more total gas and more methane gas per lb of volatile matter destroyed and fed than did any of the other lime sludge digesters.

Once again, the control digester exhibited a decrease in orthophosphate, while the lime sludge digesters showed a small release of orthophosphate. As before, the lime sludge digesters indicated a nearly complete release of magnesium into solution.

The range of operating characteristics for the digesters in the high solids run are listed in Table 9-2 and are plotted on graphs in the appendix. It is apparent that the volatile acids residual in the lime sludge digesters was unexpectedly high, especially in the higher rate (lower residence time) digesters.

This appears to be an equilibrium situation in which the methane bacteria population was limited by hydraulic washout. The concentration of soluble metals may have inhibited the growth rate of the methane bacteria. Despite the fact that the volatile acids levels were often quite high, buffering in the system prevented a significant drop in pH. For the high solids feed runs, operation was quite stable in terms of volatile matter destruction and gas production. In conventional digesters, such high levels of volatile acids would indicate incipient failure of the digestion process[14]. This is not the case for the lime sludge digesters run during high solids feed.

Table 9-2 lists some analyses run on the digested sludge. The analyses for soluble magnesium, calcium and phosphorus are averaged values near the end of each run, while the analyses for total magnesium, calcium and phosphorus are spot analyses only, done on dried sludge and converted to mg/l values. These numbers are not to be considered typical concentrations but only approximations, because not enough analyses were done to establish reliable values. This is why there are inconsistencies in comparing total calcium or magnesium in the feed sludges with those values in the digested sludges.

Looking at the values for soluble magnesium in the lime sludge digesters and comparing them to the rough values of total magnesium, one can see that most of the magnesium in the lime sludge digesters was in the soluble state. The data also show an increase in ammonia nitrogen from 120-190 mg/l in the feed sludges to 500-600 mg/l in the digester sludges for both lime and control digestion. These levels of ammonia are not considered to be in the toxic range[13].

The level of soluble sulfides in all digesters was very low. Probably most of the sulfides were precipitated as metal sulfides. The levels of grease in both the feed sludges and digested sludges can be considered only approximate values, because the number of analyses was not statistically significant.

Table 9-3 shows sludge heavy metal concentrations found in both the control digester and the lime sludge digester operated at the same 15-day SRT. Heavy metals levels except for nickel are about the same for both types of digested sludge. There is no explanation for the large difference in nickel concentrations.

The 25-day lime sludge digester which was operated at over 3 percent feed solids in the high feed solids period showed better performance than in the previous low feed solids period. Although the previous feed was only slightly lower in solids, the increase may have been enough to prevent significant washout of organisms by dilution.

Other Observations

Odor tests on the digested sludges revealed no offensive putrescible type odor from either lime sludge or control digesters in either high or low feed solids runs. In the high feed solids run, the 10-day digester sludge did have a slightly sharp odor due to the high level of volatile acids, but it was not considered offensive.

Drying tests were conducted on the digested sludges. Tests were conducted by simultaneously drying equal volumes of sludge in flat-bottomed pans both outside under direct sunlight and in the lab using a controlled temperature oven at both 80 F and 120 F. Results showed essentially the same drying rate for lime sludge and control sludge, and there was no clear effect of total solids levels on sludge drying. Dried control sludge was much darker than the dried lime sludge.

Lines and pump internals were examined at the end of the run for deposits and corrosion. The piping material was plastic (PVC) but some metal (cast iron) sections had been inserted in lines of the control and three lime sludge digesters. Results showed all lines and pump internals to be very clean. No evidence of corrosion was detected. Of course, the length of the study may have been too short to allow much deposition or corrosion to occur.

During the course of the low and high feed solids runs a main plant digester was fed a mixture of lime sludge and control primary sludge, both fed directly from primary sedimentation tanks. The mass of volatile solids fed to the digester from the control primary sedimentation tank was roughly the same as the mass of volatile solids fed from the lime sludge. During the time of monitoring, the hydraulic residence time in the digester was about 6 days, which is quite a high rate of feed. The gas composition of this digester averaged 29 percent $CO_2$, compared with an average of 37 percent for the other main plant digesters operating on primary non-chemical sludge only. The average operating characteristics were: alkalinity of 2700 mg/l; volatile acids of 840 mg/l; typical pH 7.1. The volatile matter destruction could not be precisely calculated, but was approximately 50 percent while loading in lb volatile matter per cu ft per day was 0.18. Gas production was 7.4 cu ft gas per lb volatile matter fed.

It was found that the digester was not well mixed, even though the gas recirculation system was in operation. A solids profile taken on the digester showed the following concentrations at levels below the liquid-gas interface: at approximately one-third the depth down from the interface the solids content was .62 percent TS; at approximately half way down the digester the solids content was .71 percent TS; while at the bottom draw-off the solids content was 9.6 percent TS. This shows the need for effective gas recirculation systems in digesters. Because this digester was not well

mixed, the hydraulic and solids retention times were not equal, and the 6-day hydraulic residence time was less than the solids retention time.

Other studies on digesting mixtures of lime and conventional sludges report lower $CO_2$ content in the gas, stratification in the digester, and slightly lower gas production than in conventional digestion processes[15]. A recent study of digestion of lime primary sludge with a small amount of waste activated sludge reported excellent gas quality and stable operation at both high (8-12 percent) feed TS and low (3-4 percent) feed TS[16].

High and Low Feed Solids Operations Comparisons

A comparison of the high and low feed solids runs yields the following observations:

1. Higher feed solids operation was characterized by higher carbon dioxide levels in the digester gas.

2. Higher levels of soluble calcium, magnesium and orthophosphate occurred during higher feed solids operation.

3. Lower digester pH was observed during high feed solids operation.

4. Slightly better efficiency in volatile matter destruction occurred during the high feed solids operations for SRT values up to 20 days (Fig. 9-1). Below an SRT of 20 days, the low feed solids digestion requires longer retention times than the high feed solids operation to achieve the same reduction in volatile matter.

5. Generally higher methane production per lb volatile matter destroyed (and fed) was found during the higher feed solids run.

6. Higher alkalinity and volatile acids occurred during the higher feed solids run.

7. Operation was considerably more stable when the lime digesters were fed thickened feed.

The control digester exhibited better volatile matter reduction efficiency during the higher feed solids period, but gas production (total gas as well as methane gas) per lb of volatile matter destroyed (and fed) decreased during the higher feed solids run. The methane content in the gas was slightly lower during this period as well.

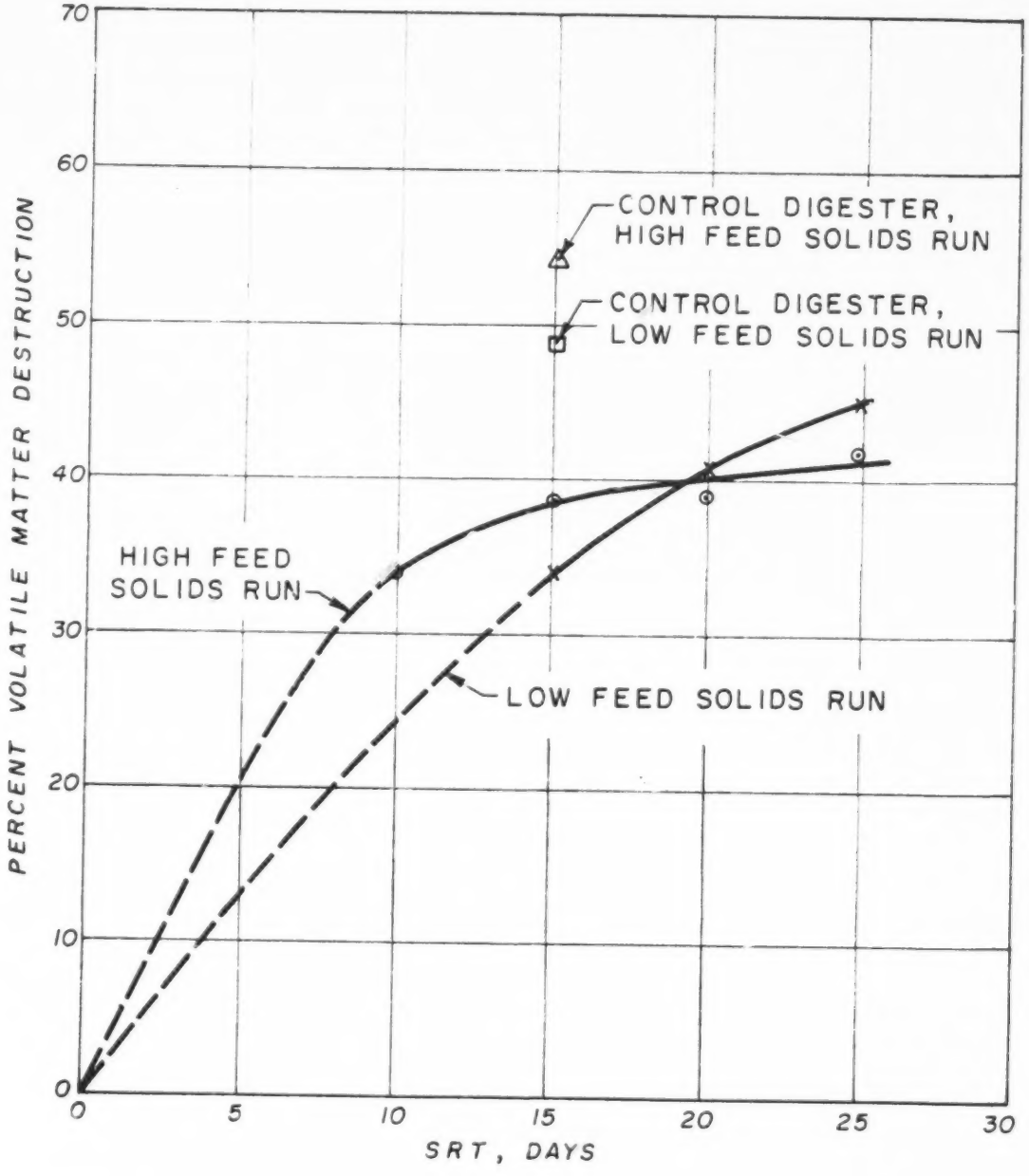



Figure 9-1

Overall Volatile Matter Destruction in Lime Digesters as a Function of Solids Retention Time

## Comparison Between Pilot Control Digester and CCCSD Main Plant Digesters

In addition to the experimental data from the pilot digester work, data from the operation of the CCCSD main plant digesters was available for the computation of digester performance parameters. Records of CCCSD digester operation (on conventional sludge) during 1971 were examined and analyzed to compare the performance of the CCCSD main plant digesters with that of the pilot control digester. The CCCSD employs three primary digesters and one secondary digester.

The pilot digester, under high (5.3 percent) feed solids operation, showed slightly less volatile matter destruction (58 percent vs. 61 percent) and unit methane production (9.8 vs. 10.7 cu ft $CH_4$ per lb VM destroyed) than did the CCCSD digesters (with 5.8 percent feed). The solids retention time, however, was significantly greater for the CCCSD digester operation (35 days) than for the pilot control digester operation (15 days). Loading rates for the CCCSD and pilot digesters were .12 and .14 (pounds of volatile matter per cubic feet of digester capacity per day) respectively. The pilot control digester gas showed lower carbon dioxide content (30 percent) than did the CCCSD digester gas (37 percent). The pilot control digester under low (2.5 percent) feed solids and a loading rate of .07 (lb VM per cu ft per day) showed less volatile matter destruction (50 percent vs. 61 percent) but higher unit methane production (12.3 vs. 10.7 cu ft $CH_4$ per lb VM destroyed) than did the CCCSD digesters.

In general, however, the performance of the pilot control digester at a 5.3 percent feed and a 15-day SRT was virtually the same as that of the CCCSD digesters under 5.8 percent feed and 35-day SRT.

The pilot control and CCCSD digester performance comparisons show that: first, digestion performance on a pilot scale digester can be considered representative of large-scale operation; and second, that the CCCSD main plant digesters at an SRT of 35 days have substantial reserve capacity.

# SECTION X
# SPECIAL STUDIES

A number of solids processing studies of an exploratory nature were made in conjunction with the main efforts. The exploratory studies included evaluation of centrate recycle to the liquid processing system, laboratory vacuum filtration, and sludge stabilization by chlorination. The results of these studies provide useful information for evaluating different processes and processing schemes.

## SIMULATED CENTRATE RECYCLE

The envisioned sludge processing scheme in the first stage of the treatment plant will include a two-stage centrifugal separation step. The calcium carbonate in the solids from the first stage will be calcined and recycled as lime to the primary treatment. Solids from the second stage, high in phosphorus, magnesium, and organic content, will be incinerated. The second stage centrate will be recycled to the head of the treatment plant. To determine if recycling the centrate will cause any serious treatment problems, a short study of centrate recycle from a simulated two-stage centrifuge system was made.

For producing a recycle centrate, a single-stage centrifuge separation was made under conditions that would remove essentially the same solids as will be removed in the two-step scheme. In earlier work, it was observed that a single-stage centrate similar in composition to a second-stage centrate can be produced by operating under conditions of high solids capture. This requires the use of a polymer, which is also in the second stage of a two-stage centrifugation process.

The total sludge production for each day was stored and centrifuged during the day shift. The centrate was pumped directly to the chemical primary precipitation tank. Consequently, the concentration of recycle centrate in the influent to the treatment plant was about three times as high during the day shift as it would be in a continuously operated process. The high concentration of recycled centrate during the day shift would be expected to accentuate any deleterious effect which might occur during continuous operation.

Due to mechanical failure of the polymer feed system the study was limited to 3½ days of satisfactory operation. For the next 1½ days the polymer was added to the centrifuge feed inlet line instead of the polymer inlet tube in an effort to continue the study after the mechanical failure. As a result of the change in location of polymer addition the amount of solids in the centrate increased substantially.

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# Table 10-1

## Effect of Recycle of Second Stage Centrate

| Process parameters | Run date, September 1972 | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | 18 | 19 | 20 | 21 | | 22 | 23 | 24 | 25 | 26 | 27 | 28 |
| Period of Centrifuge Operation | | | | | | | | | | | | |
| Start | 0930 | 1115 | 0815 | 0815 | 1200 | 0800 | | | | | | |
| Finish | 1740 | 1750 | 1615 | 1130 | 1750 | 1300 | | | | | | |
| Centrifuge Operating Conditions (P3000) | | | | | | | | | | | | |
| Feed rate, gpm | 22 | 25 | 20 | 17 | 25 | 25 | | | | | | |
| Polymer, lb per ton[a] | 1.9 | 1.4 | 1.4 | 1.5 | 3.1[b] | 4.2[b] | | | | | | |
| Centrifuge Performance[c] | | | | | | | | | | | | |
| Feed, total solids, percent | 2.6 | 3.3 | 3.7 | 2.7 | 3.1 | 2.9 | | | | | | |
| Centrate, total solids, percent | 0.3 | 0.7 | 0.7 | 0.6 | 1.5 | 1.0 | | | | | | |
| Cake, total solids, percent | 25.5 | 24.6 | 22.6 | 24.7 | 26.9 | 25.3 | | | | | | |
| Solids removed, percent | 90 | 82 | 84 | 80 | 57 | 69 | | | | | | |
| Primary Treatment Conditions | | | | | | | | | | | | |
| Influent flow, mgd | 1.2 | 1.0 | 0.8 | 1.0 | | 0.9 | 0.8 | 1.4 | 1.4 | 1.5 | 1.6 | 2.9 |
| Operating pH | 11.0 | 11.0 | 11.0 | 11.0 | | - | - | 11.0 | 11.0 | 11.0 | 11.0 | 11.0 |
| Primary Treatment Performance[d] | | | | | | | | | | | | |
| Influent SS, mg/l | 228 | 216 | 264 | 272 | | - | - | 288 | 276 | 204 | 260 | 276 |
| Effluent SS, mg/l | 27 | 28 | 33 | 27 | | - | - | 27 | 25 | 34 | 32 | 47 |
| SS removed, percent | 88 | 87 | 88 | 90 | | - | - | 91 | 91 | 83 | 88 | 83 |

[a] ICI Atlasep 3A3 polymer, lb per ton of dry solids in feed.

[b] Polymer inlet line, plugged; subsequently polymer was added to feed inlet line.

[c] Based on average of 6 to 8 grab samples per day.

[d] Results from 24 hr composite samples.

There was no apparent deterioration of the primary effluent, as shown in Table 10-1, during the 5-day period when the centrate was recycled. Visual observation of samples obtained from various depths in the primary clarifier with a Kemmerer sampler indicated no change in settling rate for the first three days. After changing the location of polymer addition there appeared to be a slight deterioration of settling rate as a result of the excess amount of solids in the centrate recycle stream.

The recovery in the centrifuge was better on the first day of operation than on the subsequent days. There was no clear indication that the difference resulted from recycled centrate. Other factors such as variations in polymer dosage and amount of cannery wastes in the sewage could have been responsible. For 2½ days of operation after the first day the total solids in the centrate and the percent solids removed were essentially constant, indicating no buildup of recycled "fines" in the sludge. The poor separation for the last 1½ days was due to the change in the location of polymer addition.

While the study was of too short duration to establish the effect of recycle with a high degree of confidence, the results did not show any deleterious effect.

## VACUUM FILTER LEAF TESTS

Filter leaf tests were conducted using primary lime sludge generated at pH 11.0, to which either anionic polymer or activated (nitrification) sludge waste solids were added. Some first stage centrate was also filtered using anionic polymer as a conditioner. All filter runs were made using an Eimco filter leaf of 0.1 sq ft area, 15 in. mercury vacuum, and Eimco filter media No. Dy-453 (yarn-spun staple). Sludge pH at time of filtration was usually below 11.0, as the pH of the sludge was observed to drop on standing, perhaps due to bacterial action.

In the first experimental series shown in Table 10-2, primary sludge was filtered at pH values of 9.4-9.5 with activated sludge solids added to give a concentration range from 0 to 3 percent of the total sludge solids. In the second experimental series, the primary sludge was filtered at pH values of 10.4-10.8 with an activated sludge solids range from 0 to 1.9 percent of the total solids. Considerably higher cake resistance values and lower filtration rates were observed for the lower pH series than for the higher pH series. The values of specific resistance within each pH range were fairly close, with no apparent relationship to the amount of activated sludge solids added.

In the next experimental series, primary sludge at pH values of 10.45-10.83 was filtered with varying amounts of polymer. With the exception of the single data point at pH 10.83 (no polymer added), there was a trend of decreasing specific resistance with increasing polymer dose up to 25 mg/l. In view of

## Table 10-2

## Vacuum Filter Leaf Studies on Primary Lime Sludge Flocculated at pH 11.0

| Sludge type and variable involved | Total solids,[a] percent | Filtration pH | Polymer dose[b] mg/l | Percent WAS solids[c] | Cake resistance[d] x $10^7$ | Filtration rate[e] | Cake TS, percent |
|---|---|---|---|---|---|---|---|
| Combined chemical sludge with activated sludge solids added at pH range of 9.39-9.55 | 3.74 | 9.39 | 0 | 0 | 9.0 | 0.7 | 28.9 |
| | 2.50 | 9.42 | 0 | 0.6 | 10.1 | 0.4 | 28.5 |
| | 3.96 | 9.55 | 0 | 1.1 | 13.7 | 0.8 | 28.7 |
| | 2.40 | 9.54 | 0 | 1.2 | 10.5 | 0.4 | 27.6 |
| | 1.97 | 9.42 | 0 | 3.0 | 10.3 | 0.4 | 27.3 |
| Combined chemical sludge with activated sludge solids added at pH range of 10.40-10.83 | 3.58 | 10.38 | 0 | 0 | 2.0 | 1.5 | 23.7 |
| | 3.59 | 10.60 | 0 | 0.4 | 1.8 | 1.9 | 21.5 |
| | 3.38 | 10.55 | 0 | 0.9 | 2.4 | 2.1 | 21.5 |
| | 3.37 | 10.50 | 0 | 1.3 | 2.0 | 1.8 | 21.4 |
| | 3.12 | 10.40 | 0 | 1.9 | 2.7 | 1.7 | 20.9 |
| Combined chemical sludge with polymer at pH range of 10.45-10.83 | 3.58 | 10.83 | 0 | 0 | 2.0 | 1.5 | 23.7 |
| | 3.20 | 10.45 | 5 | 0 | 2.8 | 1.4 | 21.8 |
| | 3.24 | 10.50 | 10 | 0 | 2.0 | 1.4 | 21.6 |
| | 2.74 | 10.50 | 15 | 0 | 1.5 | 1.4 | 23.4 |
| | 2.83 | 10.50 | 25 | 0 | 1.2 | 1.3 | 21.8 |
| First stage centrate at constant polymer dose with varying pH | 1.49 | 11.48 | 28 | 0 | 7.3 | 0.3 | 18.9 |
| | 2.22 | 10.49 | 28 | 0 | 2.3 | 0.3 | 20.1 |
| | 2.20 | 10.19 | 28 | 0 | 4.3 | 0.2 | 18.6 |
| | 2.16 | 9.59 | 28 | 0 | 6.4 | 0.2 | 18.3 |

[a] Total solids of starting sludge before any additions of polymer or waste activated sludge (WAS).

[b] ICI Atlasep 2A2 anionic polymer.

[c] Waste activated sludge solids, wt percent of total solids.

[d] Specific resistance, $sec^2$/gram, on Eimco filter leaf media No. Dy -453.

[e] Lb solids per sq ft per hr.

the results presented on the effect of sludge pH, it is probable that pH alone is the reason that the single data point at pH 10.83 without polymer has a lower specific resistance than the run at pH 10.45 with 5 ppm polymer. Despite the fairly wide range of specific resistance values, the filtration rates were all fairly close throughout the polymer dose range.

The last filtration series was conducted using first stage centrate samples at a dosage of 28 ppm polymer and varying the filtration pH from 9.6 to 11.5. Cake specific resistance values decreased as pH increased from 9.6 to 10.5, but then increased again at pH 11.5. Filtration rates were very low for all of these runs, perhaps due to the low feed solids concentration.

Using first stage centrate generated at pH 11.0, a separate experimental series consisting of three runs was made to determine the percent recovery of solids obtainable with the vacuum filter leaf apparatus. Three separate filter media were tested at a vacuum of 10 in. mercury. The results listed in Table 10-3 indicate that good solids recoveries can be achieved with vacuum filtration.

It is interesting to note that other reported values for the filterability of lime sludge (with no polymer or activated sludge added) have been much higher than the yields observed in this study. Sludge produced from a solids contact type reactor-clarifier at underflow TS of 10 to 12 percent allowed vacuum filter yields of 10 to 12 lbs per sq ft per hr[6]. These thicker sludges have much better vacuum filtration characteristics than the thinner sludges observed in this study.

## CST TESTS

A device developed by Baskerville and Gale[17], which measures capillary suction time (CST), was used to investigate the dewatering properties of combined chemical sludges. When using the CST apparatus, a sludge sample is put in a small circular tube placed vertically on filter paper. Capillary suction pressure causes filtrate to travel through the filter paper at a rate postulated to be dependent largely on the filterability of the sludge, and practically independent of the hydrostatic pressure of the sample in the vertical tube. Probes in contact with the filter paper cause the clock to start as the liquid interface reaches the first probe, and stop as the interface reaches the second probe. The time interval for the liquid to pass between the two probe reference points is related to the sludge filterability. The unit (Model 92) is manufactured by Triton Electronics, Ltd., England.

Results on a variety of chemical sludge and first stage centrate samples showed that there was no reliable correlation between capillary suction time (CST) and sludge specific resistance as measured by the filter leaf apparatus. There were also no trends in CST values with varying polymer concentrations in the sludge.

## Table 10-3

## Vacuum Filtration Tests on 1st Stage Centrate Generated at pH 11.0

| Run no. | Filter media used | Centrate TS percent | Effluent TS percent | Cake TS percent | Filtration rate, lb per sq ft per hr | Solids recovery percent w/w |
|---|---|---|---|---|---|---|
| 1 | Eimco Dy-453 spun staple | 2.38 | 0.26 | 19.3 | .45 | 94 |
| 2 | Eimco NY 332F spun fill | 2.38 | 0.42 | 20.2 | .45 | 83 |
| 3 | Eimco NY 319F multiple fill | 2.38 | 0.26 | 21.0 | .36 | 94 |

## Table 10-4

## Filtration Rates on Pepcon Processed and Unprocessed Centrates

| Centrate batch no. | Filtration rate, lb per sq ft per hr | |
|---|---|---|
| | Pepcon processed centrate | Unprocessed centrate |
| A | .08 | .08 |
| B | .12 | .11 |

## CENTRATE TREATMENT AND STABILIZATION BY CHLORINATION

A Pepcon electrolytic test facility was operated to treat centrate from the first stage lime sludge centrifuge. Basically, this unit consists of electrolytic cells, with concentric electrodes connected to a rectifier. When chlorides are present in the process stream, chlorine is produced at the electrodes and can be used for chemical oxidation of organic material. Centrate was pumped and recirculted through the annular space between the electrodes, with chlorine production and chemical oxidation occurring during this time. All tests were run at current density of 4 Kamp-hr per 1000 gal with a salt dose of 5 g/l to supply the necessary chlorides. The unit was manufactured by Pacific Engineering and Production Co. of Henderson, Nevada.

Results of processing centrate through the Pepcon unit showed that electrolytic treatment did substantially reduce biological activity and bleached the color from brownish-black to milky tan. There was also evidence of improved settling in the electrolytically processed centrate. Equal amounts of the Pepcon processed centrate and unprocessed centrate were poured into holes in the ground at opposite ends of the ATTF mobile laboratory as a qualitative test of lagooning or land disposal. The Pepcon treated centrate remained odor-free and attracted very few flies, while the untreated centrate had a pronounced odor and attracted many flies. This test was performed in the warm summer months, during which the most demanding conditions are placed on the disposal lagoons.

Pepcon-processed centrate did not show significantly improved filterability over unprocessed centrate. Filtration rates were extremely low for both processed and unprocessed centrate (Table 10-4).

Tests with the Pepcon unit were of limited duration and did not allow study of power optimization. The power cost is a significant item in considering this type of treatment. Also, salt must be added to the stream to be processed in order to facilitate chlorine production at the electrodes. If the treated centrate were concentrated by thickening prior to land disposal and the decanted fluid returned to the chemical primary, the added salt would increase total dissolved solids in the main flow stream by 60 mg/l. In light of the water reclamation objectives for this project, which limit total dissolved solids of the product water[1], this decant return could not be tolerated in the CCCSD water reclamation plant.

On the basis of these limited studies it appears that first-stage centrate can be stabilized either by chlorination or by anaerobic digestion as described in Section IX. Further engineering studies will be required to establish the practicality and economic feasibility of stabilization by chlorination.

# SECTION XI
# DISCUSSION

The studies conducted on chemical sludge production, thickening, classification and dewatering, along with the studies on digestion, provided sufficient design data to construct integrated solids processing flow sheets. Two alternative solids processing schemes have been constructed for the disposal of waste solids--one employing recalcining and incineration and the other employing digestion. In both alternative processing schemes, two stage centrifugation was used to classify and dewater (or thicken) the sludge.

## CHEMICAL PRIMARY OPERATION

Optimum primary operating pH is a function of the performances and costs of both the liquid and solids processing phases. It was shown in Section V that operation at pH 11.0 resulted in the least hardness increase across the process; this factor dictates operation at pH 11.0 for the CCCSD water reclamation plant, since minimizing hardness of the product water is desired for the purpose of industrial use. Except for this, liquid processing performance was nearly equivalent at all three operating pH levels (10.2, 11 and 11.5).

Sludge generation increased as the lime dose was elevated between pH 10.2 and 11.5. Sludge balances for four operating periods agreed exceptionally well with the solids balances constructed on the basis of changes in sewage constituents across the process. Further, sludge composition by chemical analysis was in reasonable agreement with the calculated compositions, considering the possible sources of error.

On the basis of the close agreement obtained, solids balances can be constructed for design purposes which can be expected to project realistic solids loadings.

## WET CLASSIFICATION

The extensive testing on centrifugal wet classification of combined sludge has demonstrated that the bulk of the calcium carbonate can be recovered in the cake while the undesired compounds can be rejected in the centrate. Further, this classification can be effectively accomplished with the solid bowl centrifuge over a wide range in sludge pH (pH 10.2 to 11.5). The calcium carbonate-rich cake that is produced is unusually dry, with a total solids content usually greater than 50 percent.

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## CENTRATE PROCESSING ALTERNATIVES

Wet classification by centrituge produces a centrate that contains most of the organics, magnesium hydroxide, hydroxyapatite, ferric hydroxide, and other inerts. This stream has to undergo further treatment prior to disposal. Two alternatives were considered: ( 1) dewatering by thickening and centrifugation followed by incineration and (2) anaerobic digestion of the solids followed by land disposal.

### Centrate Processing by a Second Stage Centrifuge

It was found that a solid bowl centrituge can dewater centrate from a classification centrifuge, provided an anionic polymer is fed to the machine in a dose of at least two lb per ton dry solids. The pH of operation in the primary affects the cake dryness obtained; below 11.0 a median cake total solids of 17 to 18 percent was obtained. An operating pH of above 11.0 appears to cause a deterioration in cake dryness; the median cake total solids was 12 percent. Solids recoveries of 80 percent were possible in the second stage centrifuge.

### Centrate Processing by Anaerobic Stabilization

Bench-scale studies of anaerobic digestion showed that both unthickened and thickened first stage centrate can be effectively stabilized. Volatile matter reduction is lower for centrate processing than for conventional raw sludge processing, but gas yields per unit volatile matter were the same. The sludge had an unoffensive odor in both the wet and dry states. The supernatant, however, contained virtually all of the magnesium which had been fed to the digester, as apparently all had been dissolved at the digester operating pH. This supernatant if returned to the process will cause an increase in the lime dose caused by the recycled alkalinity and magnesium.

## SOLIDS PROCESSING ALTERNATIVES

The ATTF solids processing sequence as originally proposed consisted of two stage centrifugation of the thickened primary sludge, with recalcination of the first stage cake and incineration of the second stage cake. This sequence was the basis for material balance calculations at each of the three operating pH modes.

As a second alternative, another solids processing scheme was proposed, employing land disposal of the first stage cake and digestion (with land disposal) of the second stage cake, in lieu of recalcination and incineration. A material balance flowsheet was constructed for this processing alternative, assuming a primary operating pH of 11.0 (with 14 mg/l $FeCl_3$).

Recalcination and Incineration Alternative

Three material balance flowsheets for the recalcination-incineration processing sequence were constructed (see Figures 11-1, 11-2 and 11-3) for pH modes of 10.2, 11.0 and 11.5 respectively.

Assumptions. The material balances were calculated for the first stage of the water reclamation plant at a raw sewage average dry weather flow rate of 30 mgd. The balances were based on a common set of raw wastewater qualities as listed: BOD, 216 mg/l; suspended solids (SS), 240 mg/l; phosphorus, 10 mg/l; calcium as calcium carbonate, 75 mg/l; and magnesium as calcium carbonate, 92 mg/l.

Performance of the primary at each pH level is based on the liquid processing results in Section V. Performance of the first and second stage centrifuges is based on the results presented in Sections VII and VIII.

Waste activated sludge production is based on the experimental yields determined in Section VI. All of the waste activated sludge is assumed to precipitate in the primary. Overall sludge production is estimated based on the constituent balance procedure presented in Section VI.

Incineration and recalcination performance was not tested during the project but was estimated. Dry classification, a proprietary process marketed by the Envirotech Corp., is shown in the flow sheet, but was not tested during the project. As a result no credit was given to dry classification in the balances.

It should be noted that the "organic matter" in the material balance flow sheets consists of 20 percent inert material, whereas the organic matter defined in Section IV and discussed in Sections VII and VIII consists of net "loss on ignition" material, corrected for carbon dioxide and hydrated water lost during ignition. Inerts are defined as the acid insoluble material described in the analysis procedures of Section IV.

Mass Balance Comparison. Table 11-1 summarizes the calculated results that are presented in Figures 11-1 to 11-3. At pH 11.0 less total new lime is required than at pH 11.5 or 10.2. However, the least solids are generated in the primary at pH 10.2. Thirty percent more solids are generated at pH 11.0, while 39 percent more are generated at pH 11.5 than pH 10.2. Approximately the same amount of waste solids are processed in the incinerator at each pH level. However, the amount of solids to be recalcined varies with operational pH.

A basis for judging the effectiveness of wet classification is to use the data in Fig. 11-1 to 11-3 to compute recycled solids. At pH 11.0, the centrate stream from the second stage centrifuge carries 22,059 lb/day of recycled solids. At the same time, 31,355 lb of solids are returned to the primary with the recalcined lime. This totals 53,194 lb of unwanted solids returned

FLOCCULATION CONDITIONS

289 mg/l $Ca(OH)_2$, (pH10.2), 24 mg/l $FeCl_3$, 0.25 mg/l Polymer

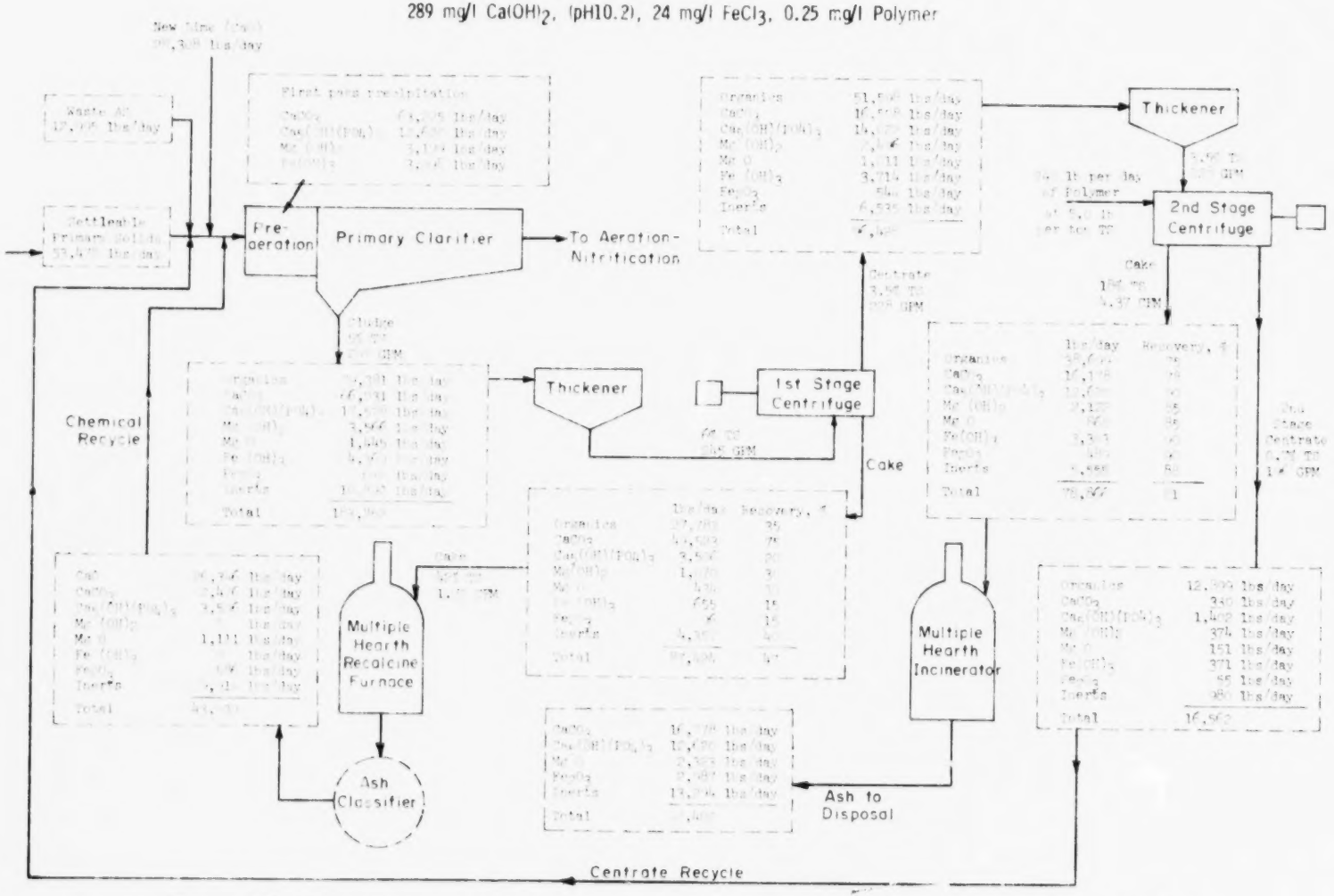


Figure 11-1 Solids Processing Balances for Incineration/Recalcination at pH 10.2

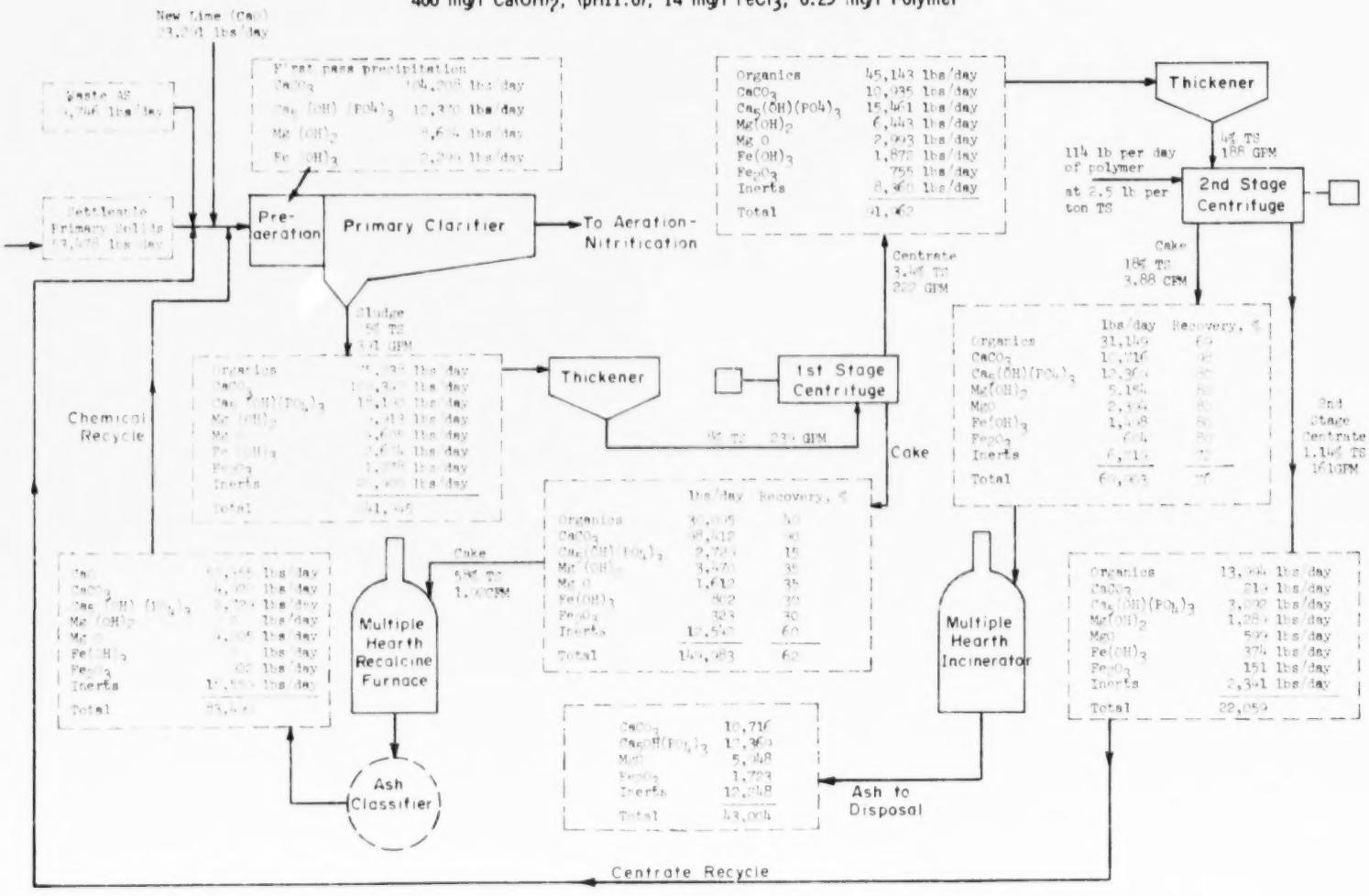


Figure 11-2 Solids Processing Balances for Incineration/Recalcination at pH 11.0

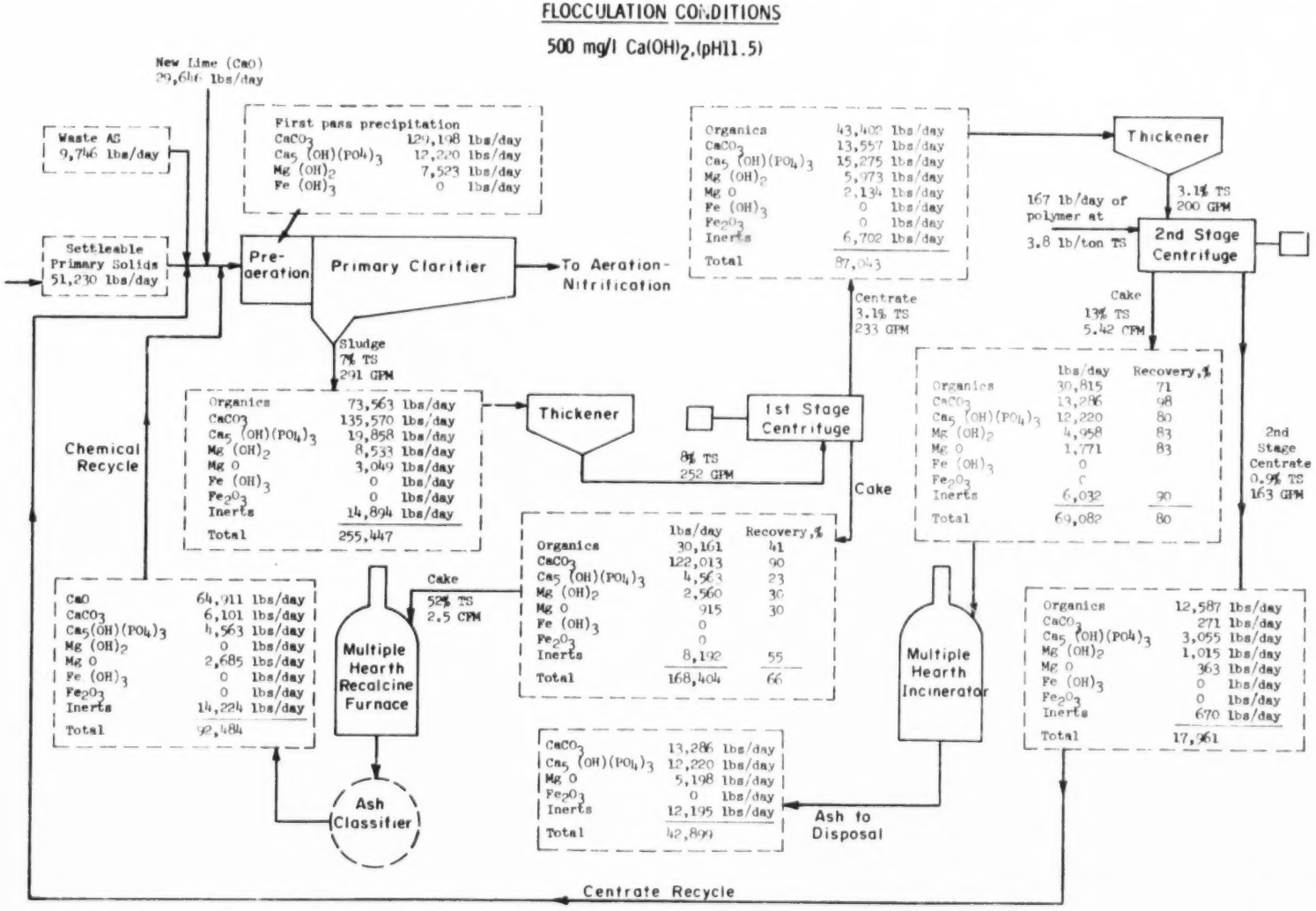


Figure 11-3 Solids Processing Balances for Incineration/Recalcination at pH 11.5

## Table 11-1

## Material Balance Comparisons at Various pH Levels

| pH | Primary sludge production | 1st Stage cake to recalcining | | Recycled solids from recalcining | 2nd Stage cake to incinerator | | Ash to disposal[a] | Makeup lime | |
|---|---|---|---|---|---|---|---|---|---|
| | lb per day | lb per day | percent TS | lb per day | lb per day | percent TS | lb per day | lb per day | Percent of total CaO dose |
| 10.2 | 184,000 | 87,000 | 42 | 44,000 | 79,000 | 18 | 47,000 | 30,000[b] | 52 |
| 11.0 | 242,000 | 150,000 | 58 | 83,000 | 70,000 | 15 | 43,000 | 25,000[b] | 31 |
| 11.5 | 255,000 | 168,000 | 52 | 92,000 | 69,000 | 13 | 43,000 | 32,000[b] | 31 |

[a]From incineration of 2nd stage cake.

[b]Basis 94% CaO.

## Table 11-2

## Chemical Costs for pH Processing Modes in Incineration/Recalcination Flow Scheme

| pH | $FeCl_3$[a] | | Primary Polymer[b] | | Centrifuge Polymer[b] | | New lime[c] | | Total daily | Total annual chemical cost |
|---|---|---|---|---|---|---|---|---|---|---|
| | lb per day | cost | lb per day | cost | lb per day | cost | lb per day | cost | cost | |
| 10.2 | 6,000[d] | $270 | 63[e] | $95 | 240[f] | $360 | 30,000 | 430 | $1155 | $422,000 |
| 11.0 | 3,500[g] | $160 | none | - | 115[h] | $170 | 25,000 | 360 | $ 690 | $252,000 |
| 11.5 | none | | none | - | 167[i] | $250 | 32,000 | 450 | $ 700 | $256,000 |

[a]Based on delivered cost of $90 per dry ton.

[b]Based on delivered cost of $1.50 per lb.

[c]Based on delivered cost of $28.69 per ton of CaO (94%).

[d]Dose 24 mg/l.

[e]Dose 0.25 mg/l

[f]Dose 5.0 lb/ton DS.

[g]Dose 14 mg/l.

[h]Dose 2.5 lb/ton DS.

[i]Dose 3.8 lb/ton DS.

to the primary clarifier, and amounts to 22 percent of the solids appearing in the underflow. Without wet classification, recycled inerts would build up to levels constituting several times the solids load that would be obtained if the solids were just incinerated and hauled away.

Chemical Costs. While a full economic comparison was beyond the scope of this study, it was possible to calculate new chemical costs for each pH level (Table 11-2).

Since pH 10.2 operation is characterized by high requirements of ferric chloride and polymer, and a higher makeup lime requirement than that for pH 11.0 operation, chemical costs are significantly higher for pH 10.2 operation ($1,155 per day) than for either pH 11.0 ($690 per day) or pH 11.5 ($700 per day) operations.

Chemical costs for pH 11.0 operation are lower than those at pH 11.5 because of both the lower makeup lime and lower centrifuge polymer requirements. However, if centrifuge polymer requirements for pH 11.5 can be reduced from the value used in the flow sheet balance (3.8 lb per ton dry solids), then pH 11.5 operation would be the least expensive of the three pH modes in terms of chemical purchase costs.

The polymer costs were scaled up from the centrifuge test work where the Sharples P600 machine was used at pH 10.2 and 11.5. The polymer dose at pH 11.0 was based on P3000 experimental data. There is reason to believe that the polymer doses required for pH 10.2 and 11.5 operations may be excessive, since comparisons of polymer requirements at pH 11.0 for both the P600 and P3000 machine showed that more polymer was required for the smaller P600 machine.

Digestion and Land Disposal Alternative

In lieu of including recalcining and incineration furnaces in the solids processing sequence, another sequence using digestion with land disposal of sludge was evaluated (Fig. 11-4).

This process includes wet classification for separation of the calcium carbonate from the rest of the sludge. This heavy calcium carbonate would pose a mixing problem in the digester if left in the sludge. The calcium carbonate-rich first stage cake can be safely disposed of on land due to its high pH. The high pH both kills pathogens and renders the sludge stable during drying[20].

The first stage centrate is then thickened in a second stage centrifuge producing a 15 percent total solids cake. A small portion of the centrate is used to dilute the cake to 8 percent for digestion, while the remaining second stage centrate is recycled to the primary. The digested solids would be disposed of in lagoons and drying beds. Supernatant from the lagoon would be pumped to the primary tank.

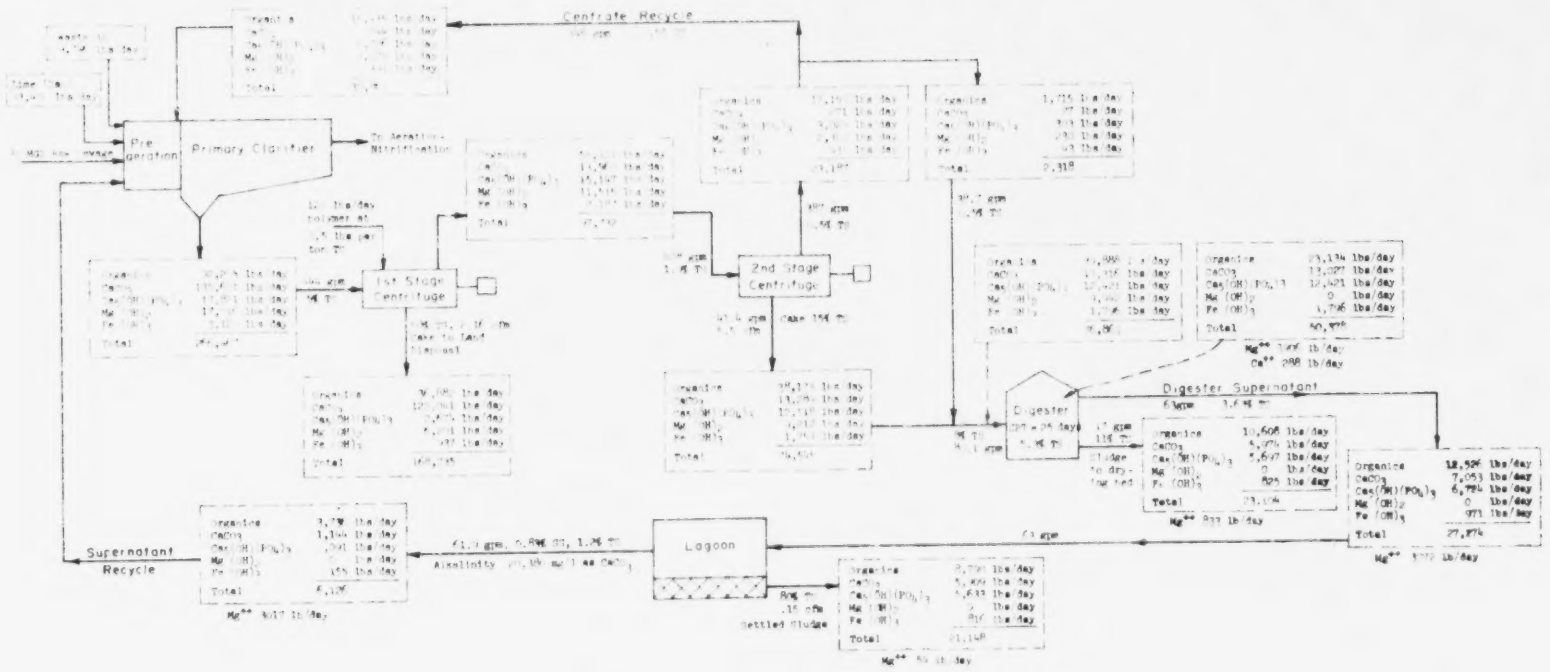


Figure 11-4

Solids Processing Balances for Digestion-Land Disposal at pH 11.0

Assumptions. Primary solids production and centrifuge operation has the same bases as the previous alternative presented.

Digester operation was based on the experimental test results presented in Section IX. These studies indicated that essentially all of the magnesium hydroxide is solubilized during the digestion process, while only a small portion of the calcium carbonate and hydroxyapatite fed to the digester is dissolved. The small incremental effect of calcium and phosphorous recycle to the primary was neglected. The main effect of returning the digester supernatant stream to the primary is the increase in lime required to react with the additional bicarbonate alkalinity and magnesium.

Digester design is based on a solids retention time of 25 days with a volatile matter destruction of 42 percent. Lagoon performance is based on the plant staff's experience with their present lagoons.

Mass Balance Comparisons. Table 11-3 summarizes the calculated results for the digestion/land disposal alternative and compares them to the recalcination/incineration alternative for pH 11.0.

The digestion alternative produces only slightly more primary sludge than does recalcination, but produces significantly greater quantities of solids for disposal. The makeup lime requirements are substantially greater without recalcination of the first stage cake. Total lime dose for the digestion alternative is greater than for the recalcination case because of the additional soluble magnesium and alkalinity being returned to the primary via the digester supernatant stream.

Chemical Costs. New chemical costs are estimated to be $635,000 per year instead of $252,000 per year for the incineration alternative. The greater cost is primarily due to the fact that no lime is recycled in the process and that the total lime requirement is greater for the digestion alternative.

Comparison to Other Digester Operations. To place the digestion alternative into perspective a comparison was made to the digester operations in a conventional plant, and an activated sludge plant as shown in Table 11-4.

For a 30 mgd plant capacity, activated sludge treatment produces more sludge (and volatile matter) to be digested than either a primary plant or an ATTF type plant using digestion to process waste solids. Based on the same raw sewage composition and flow rate, an ATTF type plant produces slightly more waste solids to digest and requires slightly more digester capacity than does a primary treatment plant. Also, because volatile matter destruction is greater for a conventional digester than for a lime sludge digester, more methane production can be expected from the conventional primary digester.

## Table 11-3

## Comparisons Between Recalcination and Digestion Solids Processing Alternatives at pH 11.0 for 30 MGD Plant

| Alternative | Primary sludge production lb per day | Lime additions[a] lb per day | Total lime[b] dose lb per day | Total solids to land fill lb per day |
|---|---|---|---|---|
| Recalcination | 242,000 | 25,000 | 76,000 | 43,000[c] |
| Digestion | 266,000 | 99,000 | 93,000 | 291,000[d] |

[a] As 94% CaO.

[b] As 100% CaO.

[c] Ash from incineration.

[d] Calcium carbonate rich cake; excludes digested solids removed by top soil company; wet basis.

## Table 11-4

## Comparisons of Large Scale Digester Operations Calculated for Three Types of Treatment Plants

| Plant type | Capacity MGD | Total sludge to digesters | | | VM to digesters | Total digester capacity | Cu ft. $CH_4$ per lb VM destroyed | Cu ft. $CH_4$ per day | SRT, days |
|---|---|---|---|---|---|---|---|---|---|
| | | lb per day | gal per day | TS, % | lb per day | cu ft. | | | |
| Primary plant[a] | 30 | 38,950 | 94,000 | 5.0 | 29,000 | 312,000[e] | 10.0 | 169,000 | 25 |
| Activated sludge plant[b] | 30 | 60,650 | 146,000 | 5.0 | 45,000 | 487,000[e] | 10.0 | 264,000 | 25 |
| ATTF type plant[c] | 30 | 76,880 | 115,000 | 8.0 | 32,000 | 386,000[e] | 11.0 | 147,000 | 25 |
| CCCSD present plant[d] | 23 | 35,496 | 73,000 | 5.8 | 27,000 | 344,000 | 10.7 | 174,000 | 35 |

[a] Based on influent SS of 240 mg/l with 65% SS removal; 75% VM in sludge solids; 58% VM destruction.

[b] Based on same data as (a) above, with additional assumption of 40% removal of influent 216 mg/l BOD across primary, and .67 lb SS produced per lb BOD fed to activated sludge (AS).

[c] Based on digestion of 1st stage centrate solids, (42% VM); pH 11.0 mode, with $FeCl_3$ and polymer addition; 42% VM destruction, with supernatant recycle to primary.

[d] Actual CCCSD operation for 1971; using three primary and one secondary digester; digester gas 37% $CO_2$; 76% VM in sludge and 61% VM destruction.

[e] Not including one secondary digester.

## SECTION XII

## ACKNOWLEDGMENTS

This investigation was a joint effort of the Central Contra Costa Sanitary District (CCCSD) and its consultant, Brown and Caldwell. Studies conducted on solids processing were part of a larger research effort conducted at CCCSD's Advanced Treatment Test Facility (ATTF). The ATTF was authorized by the CCCSD Board of Directors, who saw the need to encourage the development of tailored treatment systems suited to the purposes of reclaiming sewage for industrial use and effluent discharge. Board members are: Don L. Allan, President (1972), Parke L. Boneysteele, Director, Charles J. Gibbs, Director, Richard J. Mitchell, Director, and George A. Rustigian, Director. G. A. Horstkotte, Manager-Engineer and David G. Niles, Water Reclamation Project Manager, supervised the district's water reclamation plant expansion and research efforts.

### CCCSD STAFF

Under the direction of David G. Niles, plant superintendent, the CCCSD treatment plant staff was involved in the research program on a daily basis. Routine operation of the liquid processing system was performed entirely by the main plant's operators. Virtually all equipment installation and plant modifications were performed by the plant's maintenance staff, directed by CCCSD's mechanical engineer, Fred S. Markey. Chemical analyses of the liquid processing system were performed by the plant chemical section under the direction of Walter Fox, plant chemist. Keith Packwood had specific responsibility for scheduling the analytical program.

### BROWN AND CALDWELL STAFF

Within Brown and Caldwell, the overall plant expansion effort was supervised by Dr. David H. Caldwell, President of Brown and Caldwell. David L. Eisenhauer served as Project Manager for the design of the plant expansion and also supervised the design of the Advanced Treatment Test Facility. Dr. Denny S. Parker directed process research at the ATTF. On the Research staff at the ATTF were Fred Zadick, environmental/chemical engineer, James Tyler, supervising chemist, and Kenneth Train, chemist/chemical engineer.

### OTHER CONTRIBUTORS

Environmental Quality Analysts, a division of Brown and Caldwell, performed the detailed sludge analyses.

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The Sharples Division of the Pennwalt Corporation participated in the research on lime sludge centrifugation. Ken Kyte of the Warminster office was particularly helpful, as he was able to utilize his experience on centrifugation of lime sludges obtained at several other plants. Clem O'Donnell and Jim McCormick, from the San Francisco office of Pennwalt Corporation, assisted in many of the experimental runs.

A number of equipment manufacturers loaned equipment which significantly expanded the scope and capability of the research program. The Sharples and Wallace and Tiernan Divisions of the Pennwalt Corporation loaned centrifuges, sludge pumps, and a variable speed chemical feed pump. The Dow Chemical Company loaned chemical injectors, a variable speed chemical feed pump, and a Lightnin mixer for mixing polymer solutions. The Komline-Sanderson Corporation loaned a dissolved air flotation unit equipped with influent and chemical feed pumps. The Bird Machinery Corporation loaned a centrifuge and participated in several days of testing. The Pacific Engineering and Production Company of Nevada loaned the Pepcon chlorine oxidation system and operated it for a week.

This project was funded from several sources. The California State Water Resources Control Board and U.S. Environmental Protection Agency each contributed a grant to the district. The district contributed an equal amount. Project officers for the State Board and EPA were Kurt Wassermann and Robert Dean, respectively.

## REPORT PREPARATION

This report was written by Brown and Caldwell. Principal authors were Denny S. Parker, Fred Zadick and Kenneth Train.

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## SECTION XIV

## GLOSSARY

| | |
|---|---|
| amp | ampere(s) |
| AS | activated sludge |
| C | degrees Centigrade |
| $C_c$ | constituent in total solids of centrifuge cake, weight percent |
| $C_t$ | total solids in centrifuge cake, weight percent |
| cfm | cubic feet per minute |
| cu ft | cubic feet |
| DS | dry solids |
| $c_t$ | total solids in centrifuge centrate, weight percent |
| F | degrees Farenheit |
| $f_c$ | constituent in total solids of centrifuge feed, weight percent |
| $f_t$ | total solids in centrifuge feed, weight percent |
| g | gravitational force per pound mass, feet per (second)$^2$ |
| gal | gallons |
| gpd/sf | gallons per day per square foot |
| gpm | gallons per minute |
| in. | inches |
| lb | pound |
| LOI | loss on ignition |
| mg/l | milligrams per liter |
| MLVSS | mixed liquor volatile suspended solids |
| M/N | methanol to nitrogen (mass ratio) |
| $NH_3$-N | ammonia nitrogen |
| ppm | parts per million |
| $R_c$ | weight percent recovery of a given constituent |
| $R_t$ | weight percent recovery of total solids |
| △ RPM | difference in revolutions per minute. For centrifuge, △ RPM is bowl speed minus auger speed. |
| sq ft | square feet |
| sq in. | square inches |
| SRT | solids retention time |
| SS | suspended solids |
| SVI | Sludge Volume Index |
| TOC | total organic carbon |
| TS | total solids |
| TSS | total suspended solids |
| VM | volatile matter, interchangeable with volatile solids in this report |
| VS | volatile solids (or volatile matter) |
| VSS | volatile suspended solids |
| WAS | waste activated sludge |
| wt | weight |

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# SECTION XV. APPENDIX

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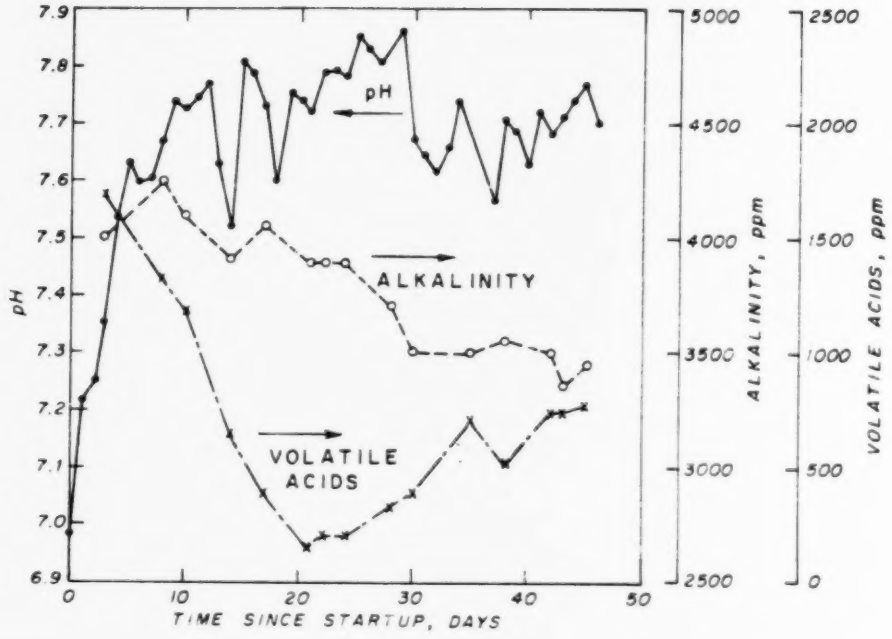


Figure A-1
Lime Digester Operation During Low Feed Solids Period for SRT of 15 Days

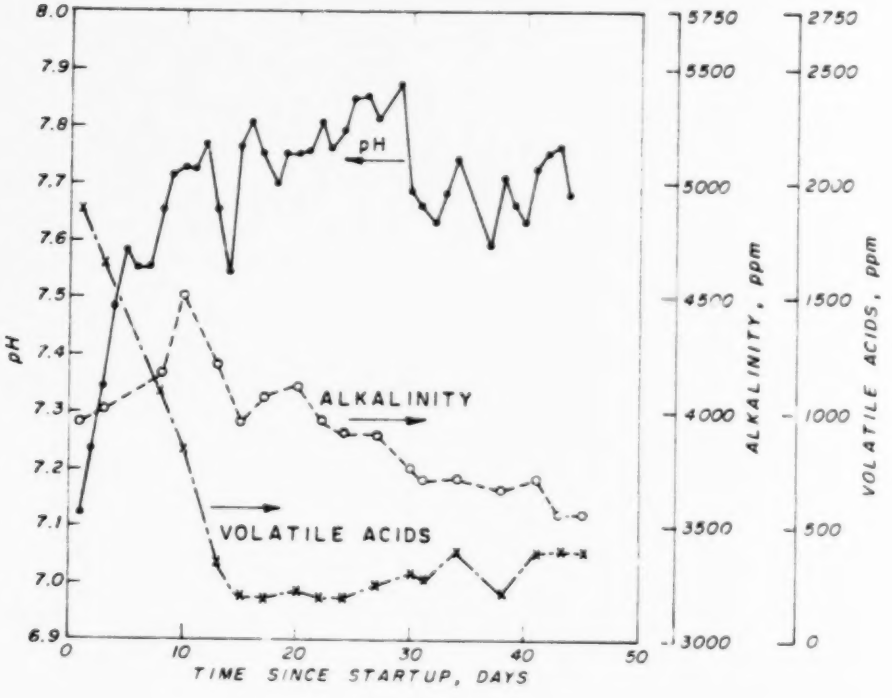


Figure A-2
Lime Digester Operation During Low Feed Solids Period for SRT of 20 Days

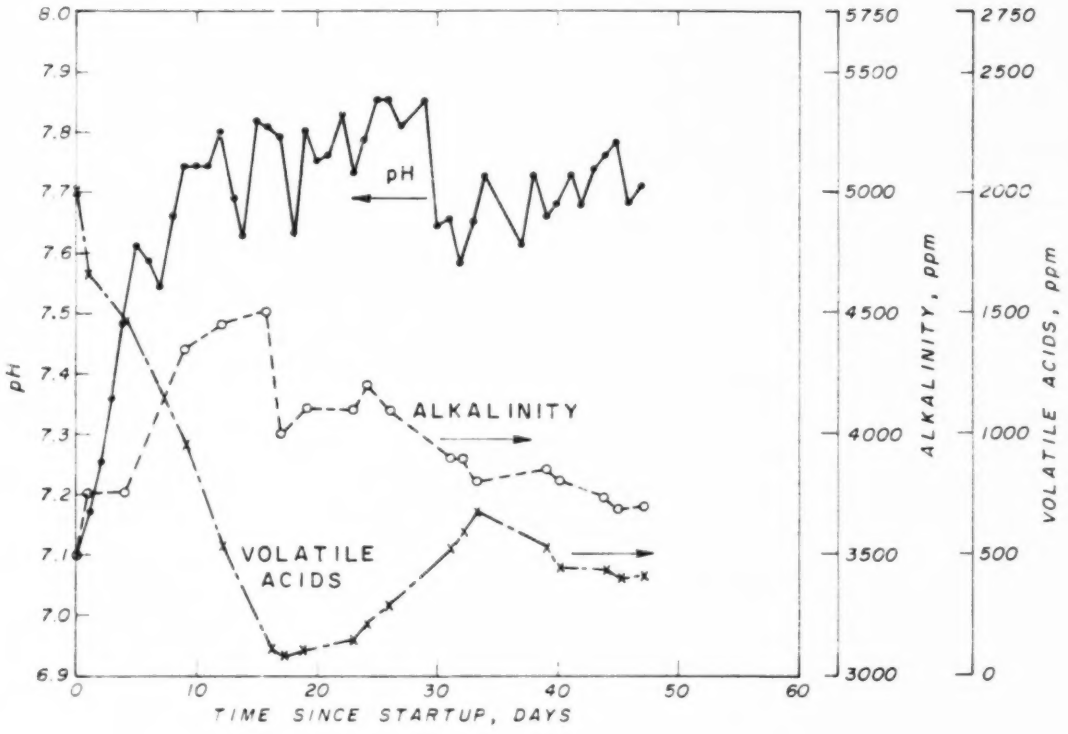


Figure A-3

Lime Digester Operation During Low Feed Solids Period
for SRT of 25 Days

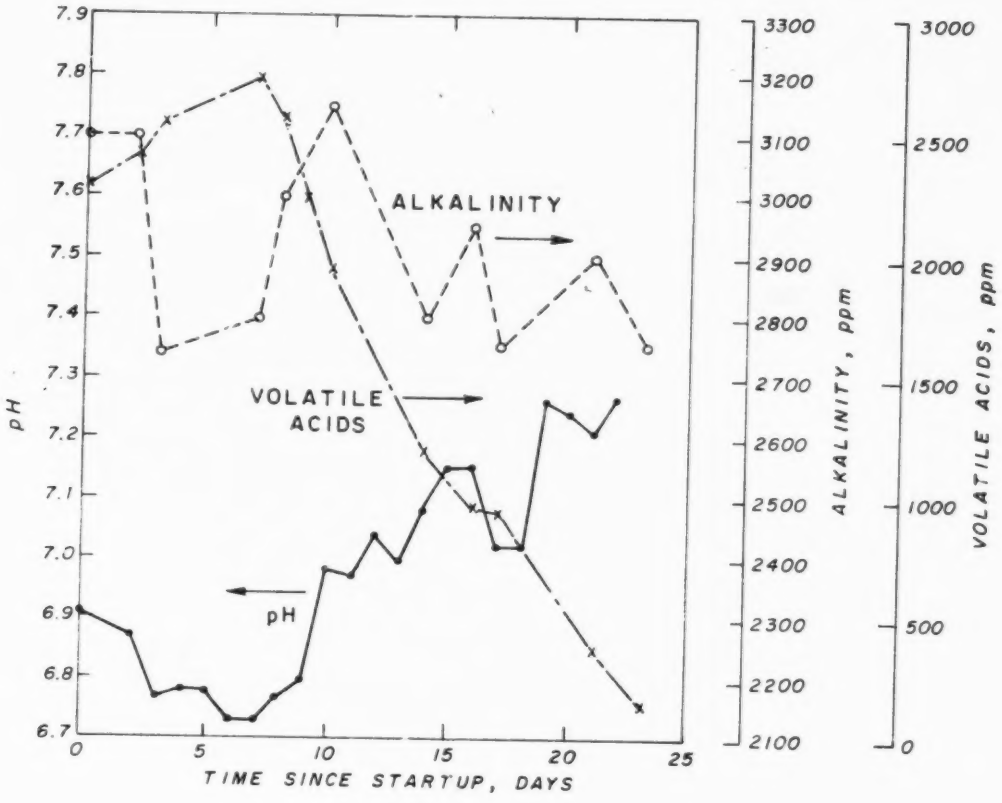


Figure A-4

Control Digester Operation During Low Feed Solids Period
for SRT of 15 Days

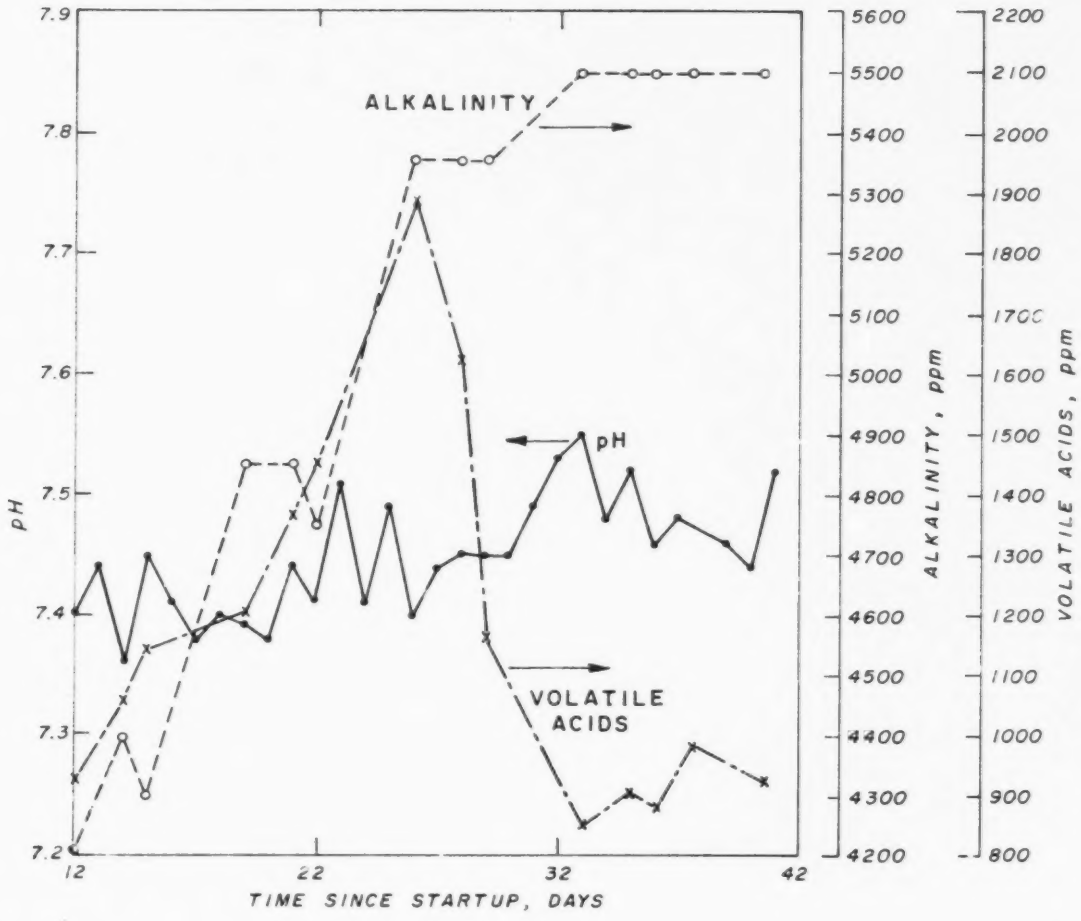


Figure A-5

Lime Digester Operation During High Feed Solids Period for SRT of 10 Days

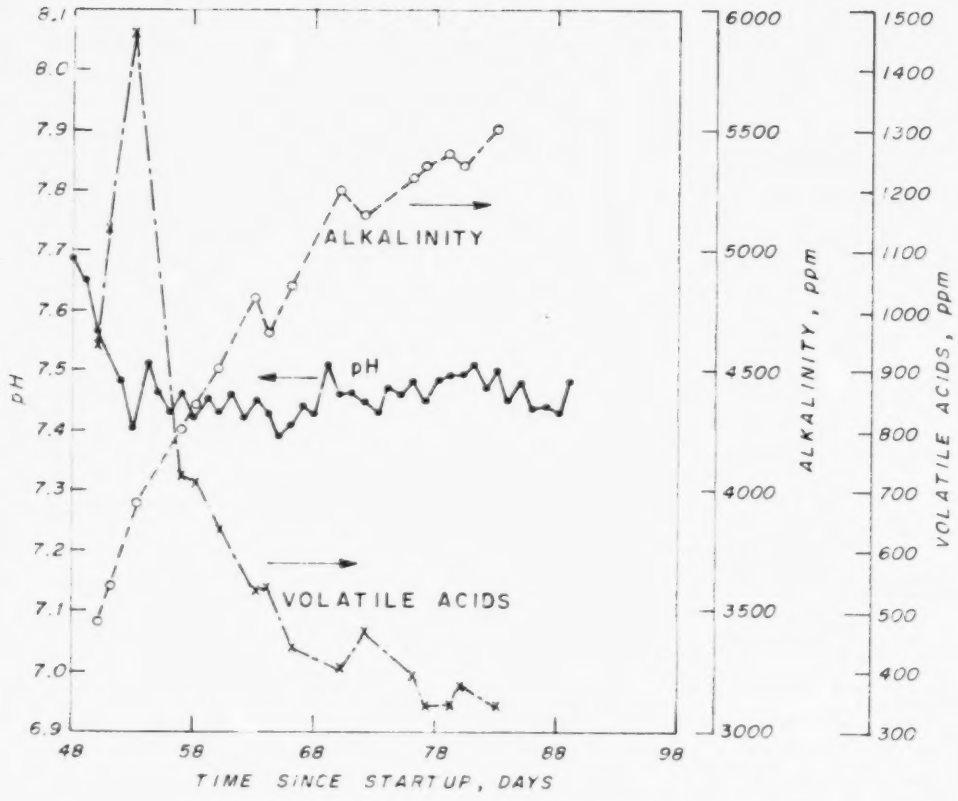


Figure A-6

Lime Digester Operation During High Feed Solids Period
for SRT of 15 Days

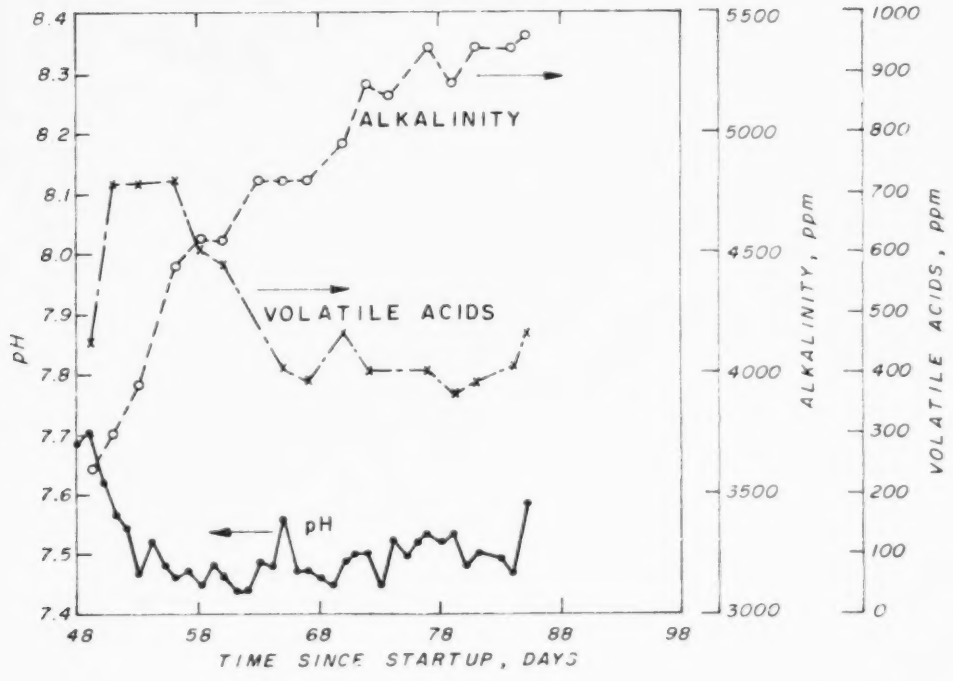


Figure A-7

Lime Digester Operation During High Feed Solids Period
for SRT of 20 Days

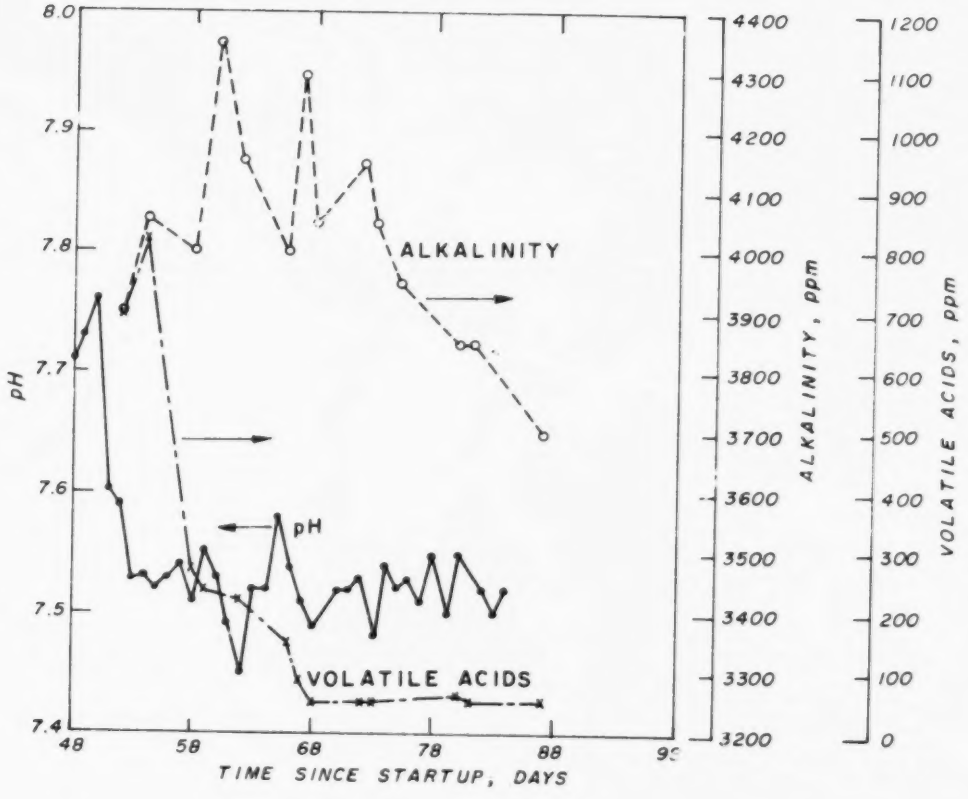


Figure A-8

Lime Digester Operation During High Feed Solids Period for SRT of 25 Days

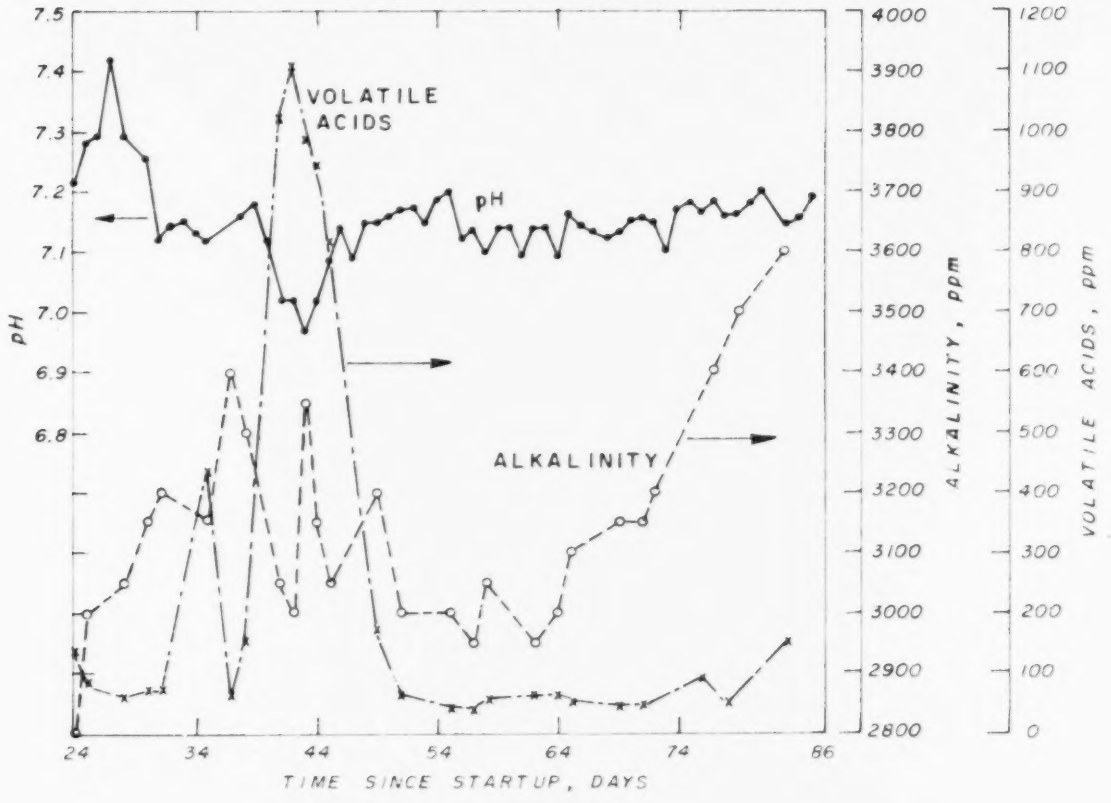


Figure A-9

Control Digester Operation During High Feed Solids Period for SRT of 15 Days

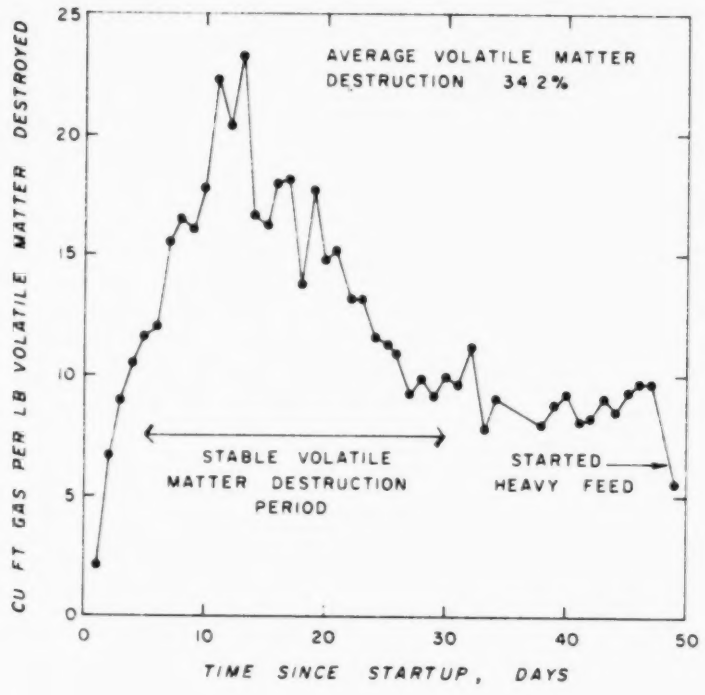


Figure A-10

Unit Gas Production for Lime Digester During Low Feed Solids Period, SRT of 15 Days

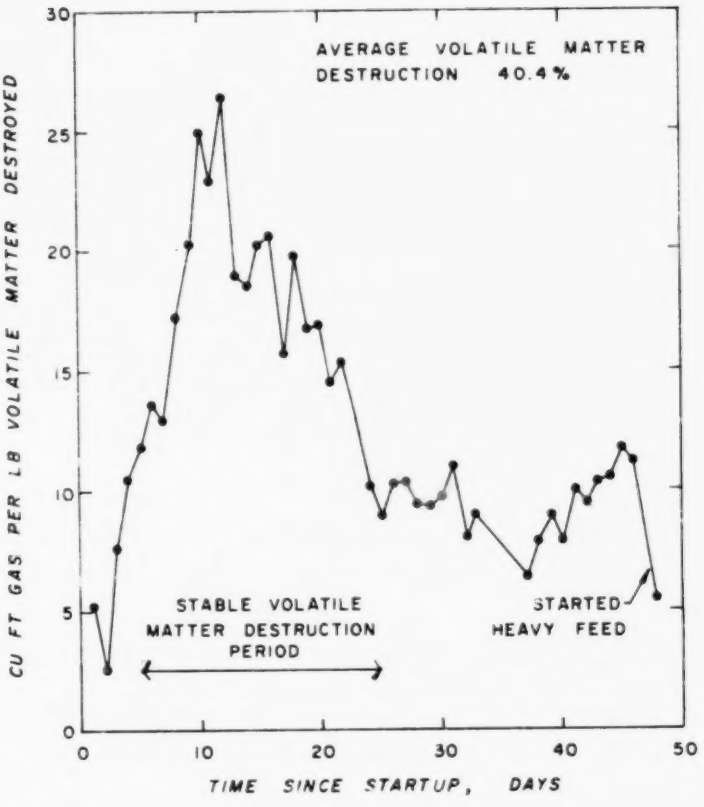


Figure A-11

Unit Gas Production for Lime Digester During Low Feed Solids Period, SRT of 20 Days

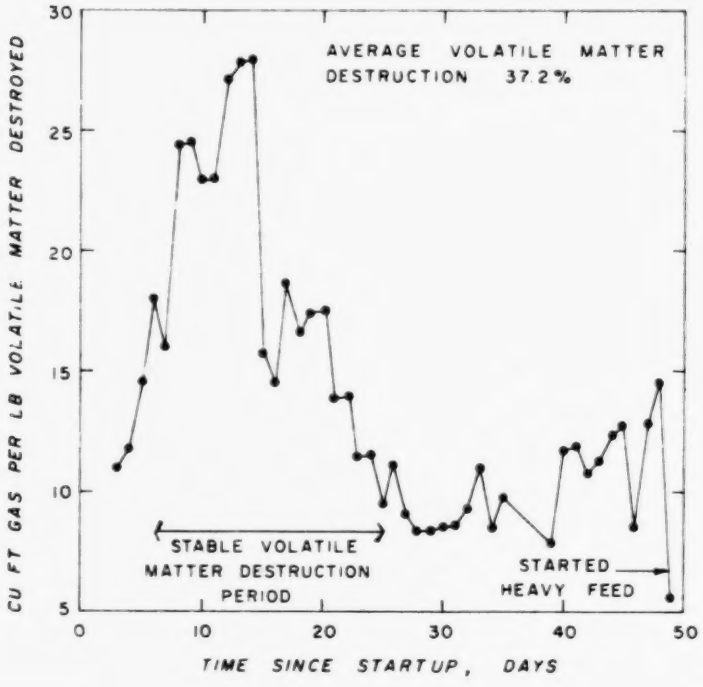


Figure A-12

Unit Gas Production for Lime Digester During Low Feed Solids Period, SRT of 25 Days

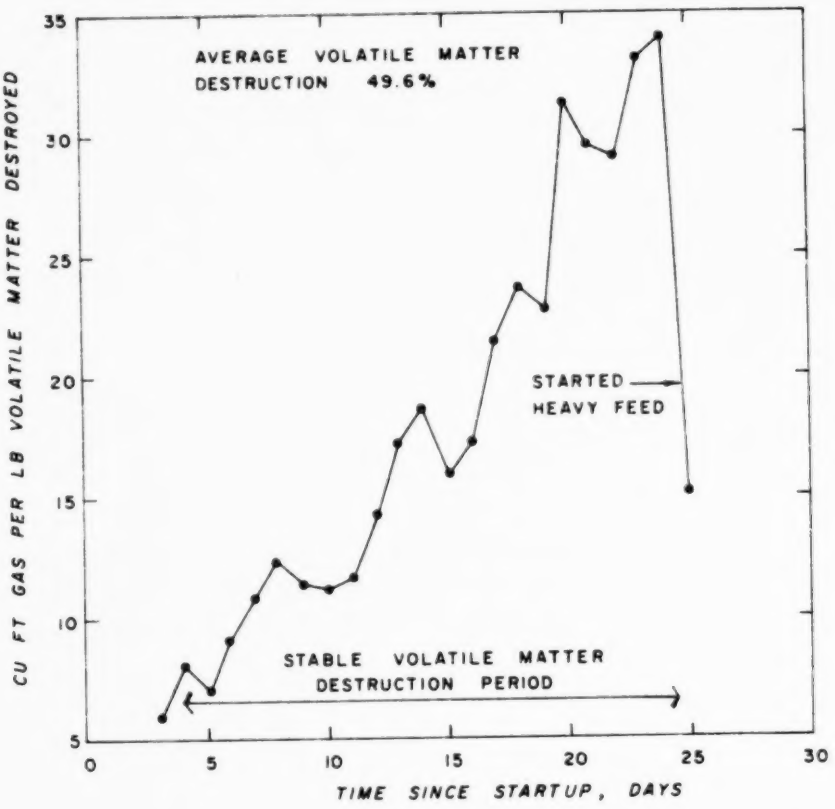


Figure A-13

Unit Gas Production for Control Digester During Low Feed Solids Period, SRT of 15 Days

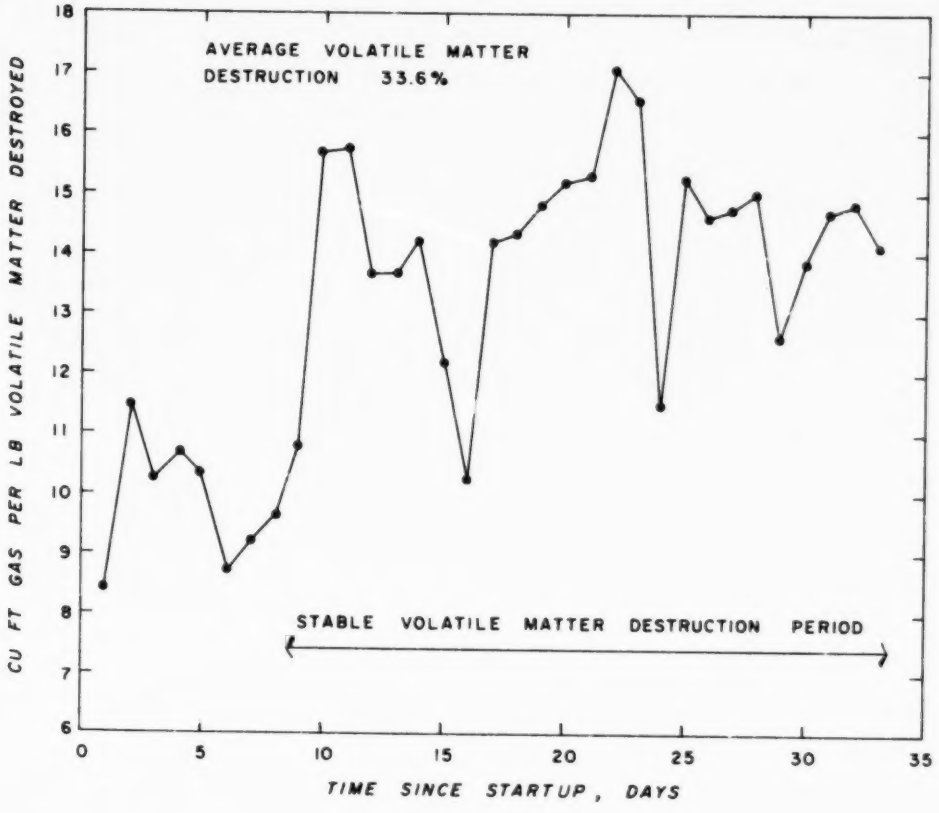


Figure A-14

Unit Gas Production for Lime Digester During High Feed Solids Period, SRT of 10 Days

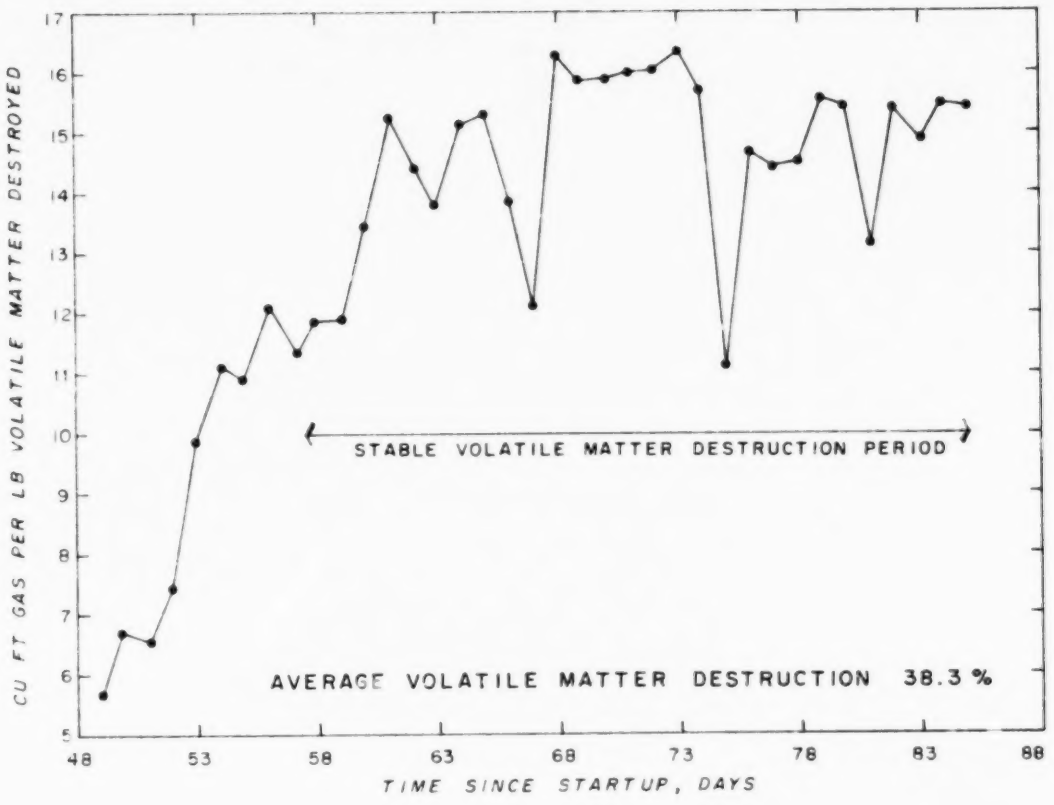


Figure A-15

Unit Gas Production for Lime Digester During High Feed Solids Period, SRT of 15 Days

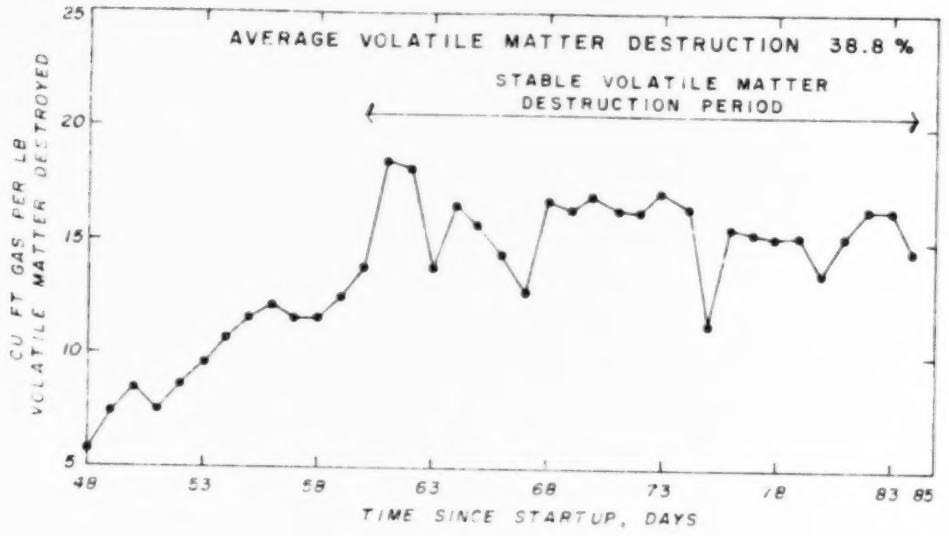


Figure A-16

Unit Gas Production for Lime Digester During High Feed Solids Period, SRT of 20 Days

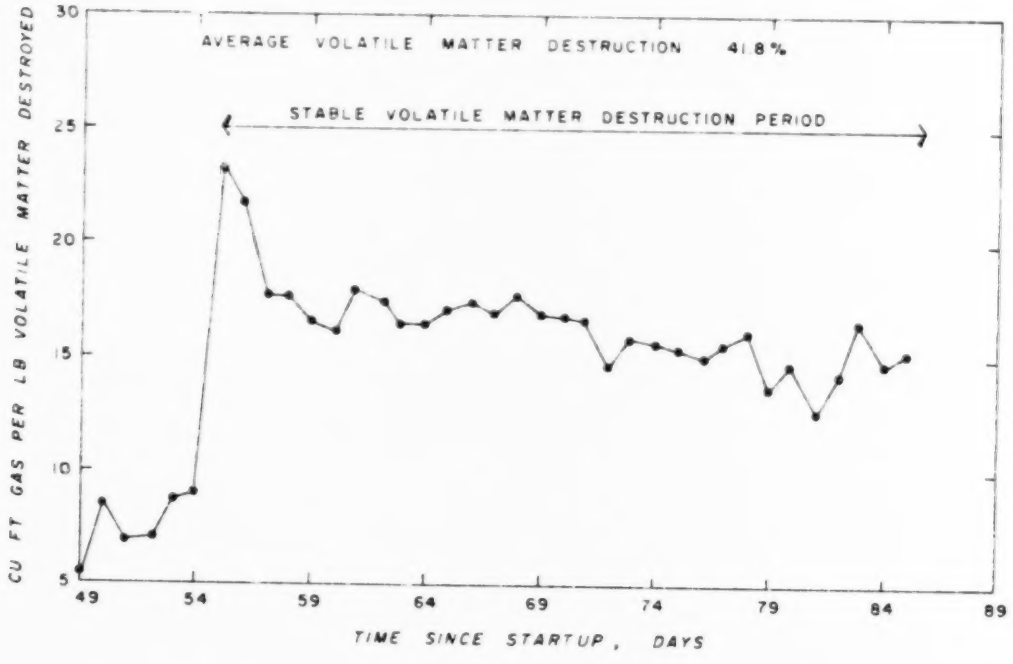


Figure A-17

Unit Gas Production for Lime Digester During High Feed Solids Period, SRT of 25 Days

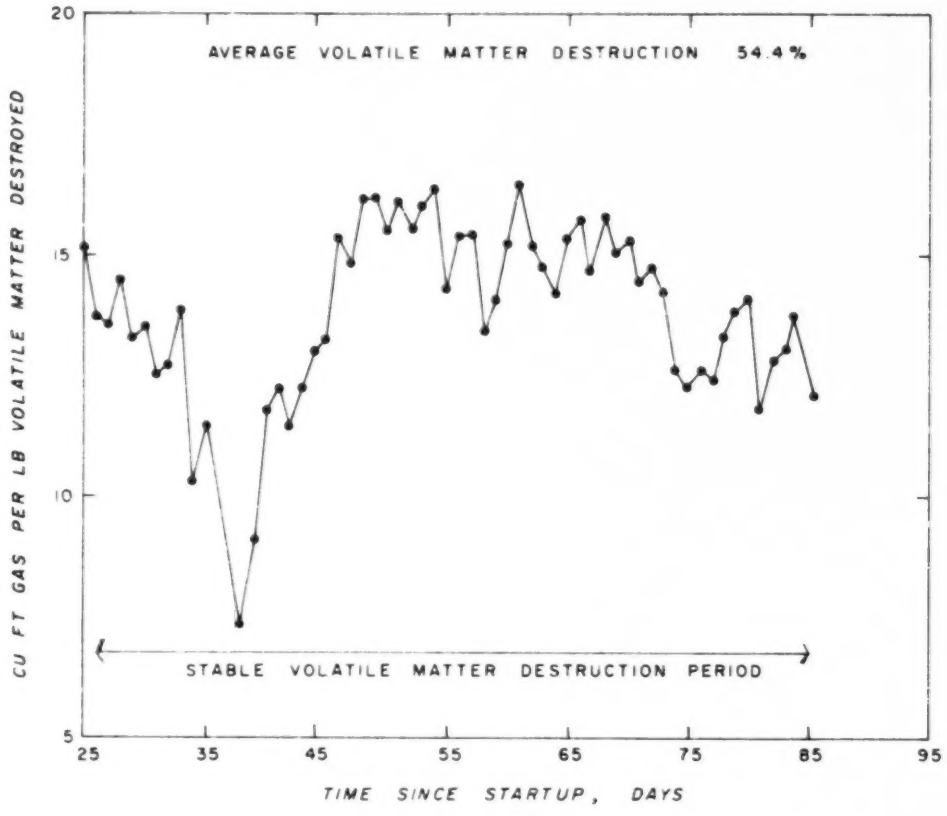


Figure A-18

Unit Gas Production for Control Digester During High Feed Solids Period, SRT of 15 Days



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